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VOL. 41 NO. 376 APRIL 1955

THE INSTITUTE OF PETROLEUM

AN Ordinary General Meeting of the Institute of Petroleum was held at 26 Portland Place, London, W.1, on 8 December 1954, the Chair being taken by the President, Lt.-Col. S. J. M. Auld, O.B.E., M.C., D.Sc.

The Editor read the minutes of the previous meeting, which were confirmed and signed as a correct record. He also announced the list of members elected since the previous meeting, and gave notice of the date of the next meeting.

The Chairman, introducing the authors of the paper to be presented, said: We have, in the past, been indebted to the Fuel Research Station for many interesting and important papers, and consequently I think we have all been looking forward to this evening to hear the communication from Dr Hall and Mr Taylor, two of the senior members of the staff of that station.

The Fischer–Tropsch process has held our attention in numerous ways and over a good many years, and its continuance of interest is evidenced by the attendance tonight; a good many of you, I know, being our guests from the Institution of Chemical Engineers, and to you guests, I offer a sincere welcome and the hope that you will come again.

I will now ask Dr Hall to present the paper on behalf of himself and his colleague.

DESIGN AND OPERATION OF A FLUID CATALYST PILOT PLANT FOR FISCHER–TROPSCH SYNTHESIS

By C. C. HALL* and A. H. TAYLOR*

INTRODUCTION

THE development of improved processes for the synthesis of liquid fuels and chemical products from carbon monoxide and hydrogen necessitates research along two general lines. One involves the study of catalyst activity and selectivity and the effect of the process variables (temperature, pressure, gas composition, etc.) on the rate and course of the reaction. This type of work can, in general, be carried out in simple, small-scale tubular reactors containing a bed of catalyst granules or pellets. The other line is concerned with the development of a suitable reaction system for carrying out the exothermic and highly temperature-sensitive reaction on a commercial scale. Although a valuable preliminary assessment of the suitability of a particular system can be obtained in equipment with a reaction space of, say, 1 to 2 litres, it is ultimately essential to build and operate a pilot plant of sufficient size to provide a more reliable indication of the probable performance obtainable in a full-scale plant.

In the work on the Fischer–Tropsch synthesis at the Fuel Research Station, two reaction systems, which appear to have advantages over the original fixed-bed system, have been studied. Work on fluidized catalysts (originally applied to this type of process in the U.S.A.)[1] was started in 1947, and the results of experimental work in small-scale (1-inch and ½-inch dia) reactors have been published.[2–5] Study of the liquid-phase ("slurry") system was initiated somewhat later, and the essential features of this system were compared with those of the fluid-catalyst and fixed-catalyst systems (when all three are operated under the same conditions) in a paper read before the Institute in March 1952.[4]

In 1949 it was decided to erect a pilot plant for the study of various process techniques on a scale of 1 to 2 barrels of products per day. At that time, the fluid-catalyst technique appeared to be particularly promising, and operation of small units had provided sufficient information to permit the design of a larger unit. It was decided, therefore, to build the pilot plant for operation, in the first instance, with fluidized catalyst. The potential advantages of the slurry system were fully recognized, but further work on a smaller scale appeared desirable before constructing a pilot-scale reactor for this process. The general plan was the installation of the necessary equipment for purification, measurement, and compression of 2000 cu. ft/hr of synthesis gas and for the handling of a product output up to 80 gal per day. A change in the particular synthesis technique employed would, therefore, require only a change in the reaction vessel and certain items of equipment directly associated with it—a comparatively minor proportion of the whole plant in terms of cost and labour.

By the time the plant was ready for operation (September 1952) it had become apparent that the dense-phase fluidized-catalyst technique possessed certain limitations and disadvantages and was probably not the most suitable one to employ where coal is the basic raw material. It appeared, furthermore, that the fluid-catalyst synthesis plant built for Carthage Hydrocol Inc. at Brownsville, Texas, had run into serious difficulties, certain of which, apparently, were associated with the synthesis reactors. This could be interpreted as evidence that

* Fuel Research Station, D.S.I.R.

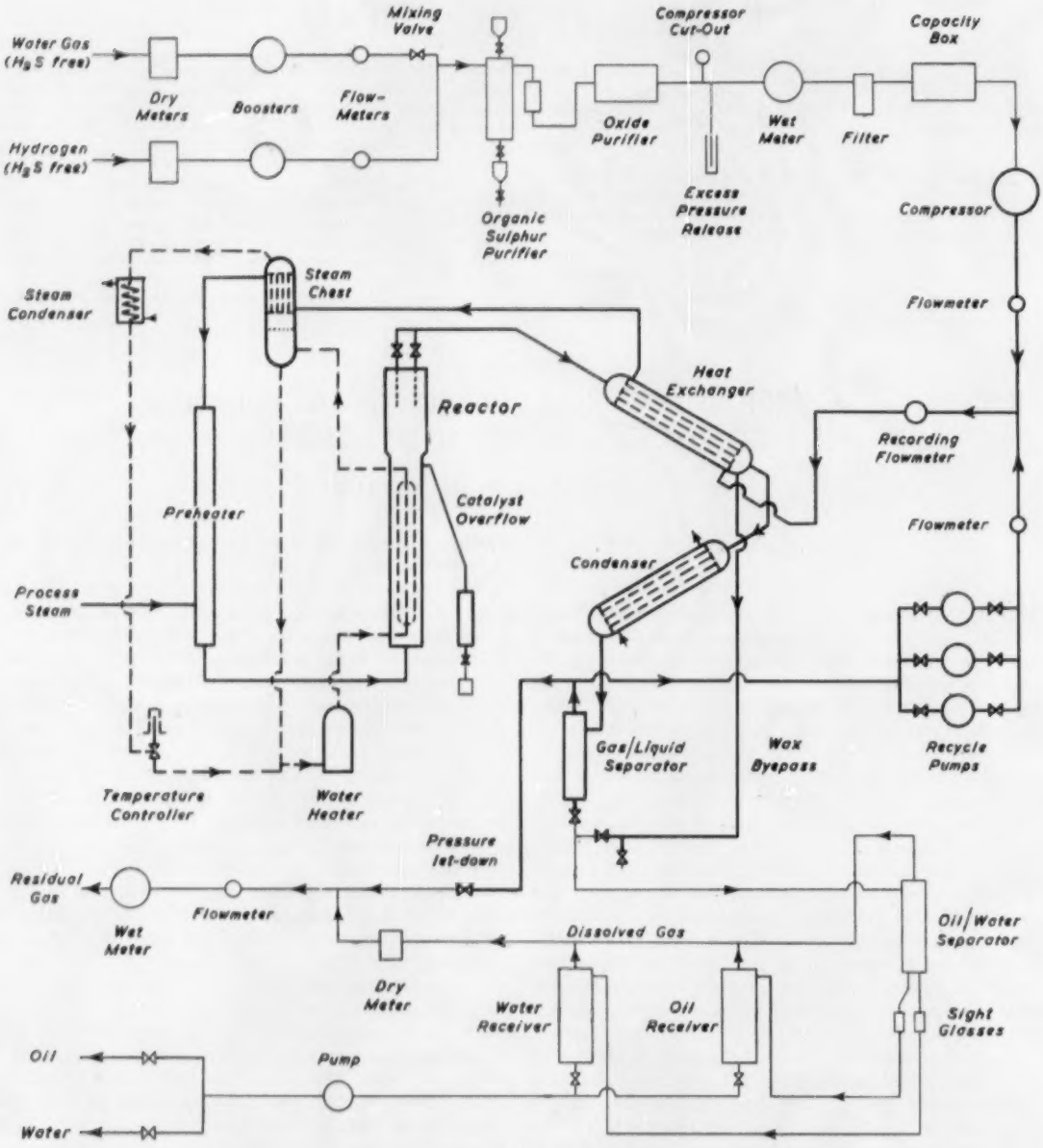


FIG 1

PILOT SCALE FLUIDIZED CATALYST HYDROCARBON SYNTHESIS PLANT. FLOW DIAGRAM

pilot plants of a similar size to that built at Greenwich had proved to be an unreliable basis for the design of full-scale plants. As, however, no details of the design or performance of either the pilot plant or the full-scale plant built in the U.S.A. have been published, it was considered to be desirable to obtain some first-hand knowledge of the fluid-catalyst synthesis process when operated on a pilot-plant scale.

In this paper a description of the plant is given, and the results obtained in two test runs of 14 and 30 days duration are reported and discussed.

DESCRIPTION OF PLANT

The plant was designed for a throughput of fresh feed gas up to 2500 cu. ft/hr (s.t.p.), operating pressures in the range 20 to 50 atm, and reaction temperatures up to 340° C. Parts of the plant coming into contact with condensing products were made in 18/8/1 stainless steel, all other parts in mild steel. The feed gas was obtained by blending blue water-gas and hydrogen from the steam–iron process. Both gases were available on the site in adequate quantities; they were purified to remove hydrogen sulphide and were piped to the plant building. A simplified flow sheet for the system is shown in Fig 1. Design and operational details of the main items of equipment are given below.

Reaction Vessel

The constructional details of the fluid-catalyst reactor are shown in Fig 2. The reaction space is $7\frac{7}{8}$ inches × 7 ft 6 inches high, the effective cross-sectional area being 0·257 sq. ft. and the volume 1·95 cu. ft. This is the maximum volume which the catalyst can occupy, the limit being set by the catalyst-overflow line. The cross-sectional area of the upper part of the reactor (the "disengaging space") is 0·52 sq. ft. A general view of the reactor is shown in Fig 3.

Gas distribution inside the reactor is obtained by a sintered bronze (Porosint) disk $\frac{1}{8}$ inch thick, 12·5 μ pore size, in the base. Its effective area is about 0·15 sq. ft. This replaced the shallow bed of steel balls which proved unsatisfactory in the trial run.

Two filter units are fitted at the top to prevent catalyst being carried over into the condensing system. Those fitted initially were "star" filters, one of sintered stainless steel and one of sintered bronze, approximately 5 μ pore size. Each had a filter area of 0·75 sq. ft. When, at the end of Test Run 1, some of the joints of the stainless steel filter were found to have opened, both star filters were discarded. They were replaced by the two tubular units shown in Fig 2, which are sintered bronze, 12·5 μ pore size, each unit having an area of 1·95 sq. ft. Provision is made for intermittent blow-back with purge gas.

The heat of reaction is removed by the circulation of water under pressure through a tube bundle in the reactor. The system is shown in Fig 5. Steam generated separates in a steam chest at high level,

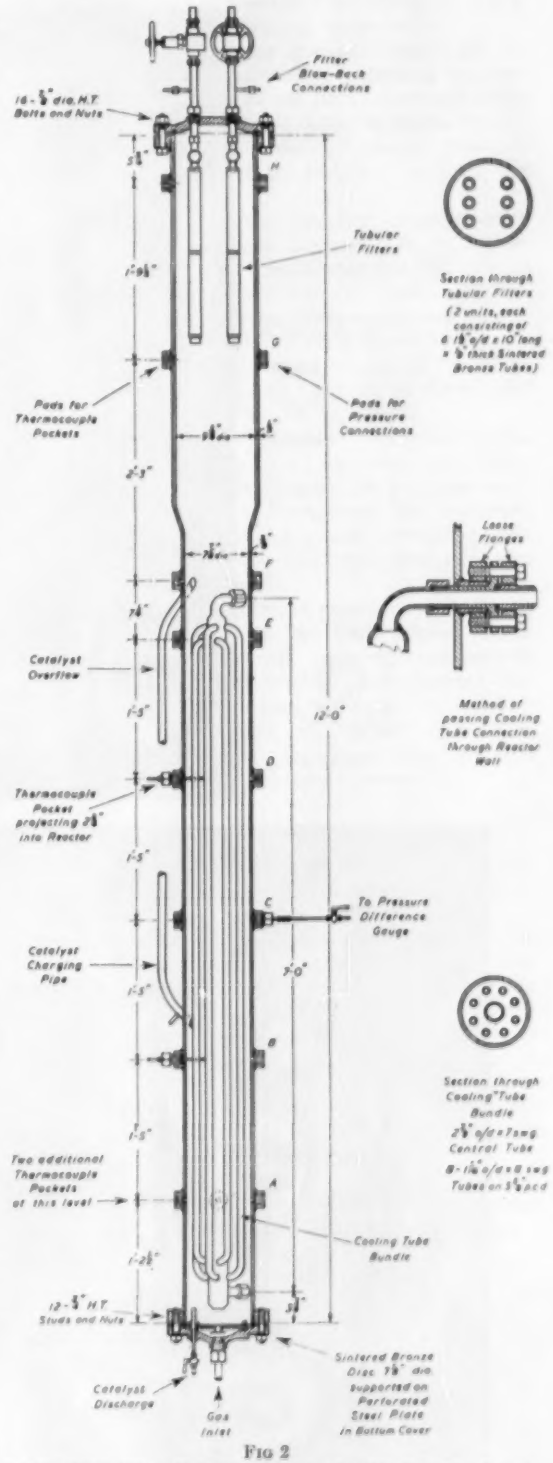

Filter Blow-Back Connections
16 - ⅝" dia. H.T. Bolts and Nuts
Tubular Filters
Section through Tubular Filters
(2 units, each consisting of 6 1⅝" o/d x 10" long x ⅛" thick Sintered Bronze Tubes)
Pads for Thermocouple Pockets
Pads for Pressure Connections
Loose Flanges
Catalyst Overflow
Method of passing Cooling Tube Connection through Reactor Wall
Thermocouple Pocket projecting 2⅛" into Reactor
To Pressure Difference Gauge
Catalyst Charging Pipe
Section through Cooling Tube Bundle
2⅛" o/d x 7 swg Central Tube
8 - 1⅛" o/d x 8 swg Tubes on 3½" pcd
Two additional Thermocouple Pockets at this level
Cooling Tube Bundle
12 - ⅝" H.T. Studs and Nuts
Sintered Bronze Disc 7⅞" dia. supported on Perforated Steel Plate in Bottom Cover
Catalyst Discharge
Gas Inlet

FIG 2

PILOT SCALE FLUIDIZED CATALYST HYDROCARBON SYNTHESIS PLANT REACTOR

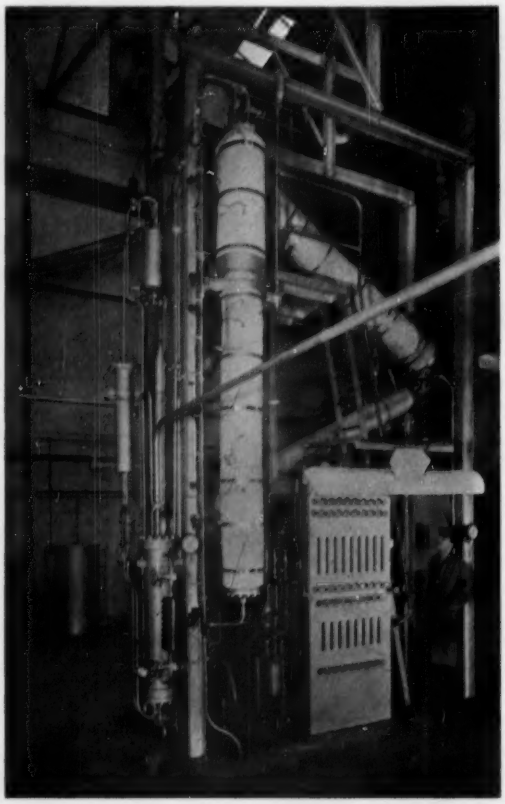
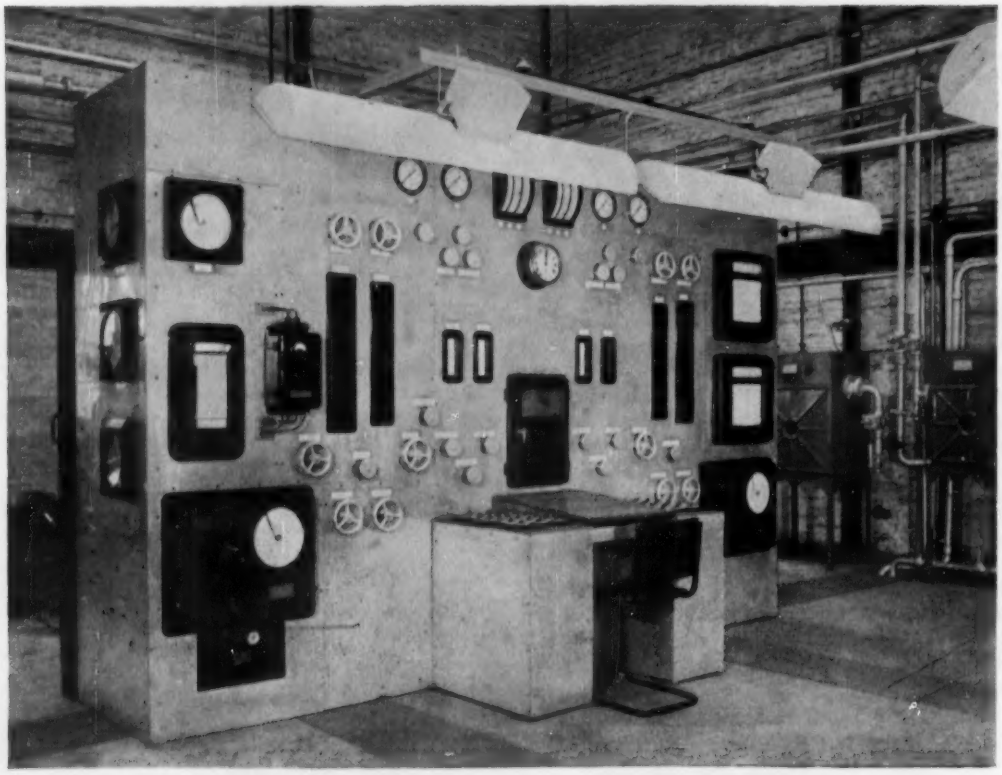


FIG 3
VIEW OF THE PLANT SHOWING THE REACTOR AND THE DIFFERENTIAL PANEL

FIG 4
THE MAIN INSTRUMENT AND CONTROL PANEL



and the water returns to the tube bundle via a vessel fitted with electric heaters. Some of the steam is condensed in providing part of the preheat to the feed gas, the remainder is condensed in a water-cooled coil, the rate through which is governed by a "dead-weight" valve, set by external weights to maintain any desired steam pressure. The temperature of the effluent cooling water is continuously recorded. The system is a closed one, but provision is made for pumping in water during operation when necessary. Circulation is by thermo-syphon. Pressures up to

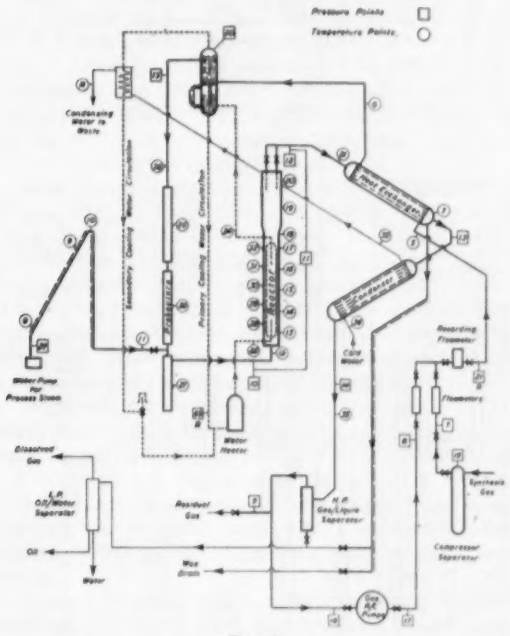


FIG 5

PRESSURE AND TEMPERATURE POINTS

85 atm (= 300° C) were used during the runs described. The tube bundle has an external cooling surface of 19·0 sq. ft.

The method shown of connecting the tube bundle to the reactor was adopted to allow easy removal if it was desired to alter the cooling surface.

Main Instrument and Control Panel

All gas flows, temperatures, pressures, etc., which are important for the operation of the plant, are measured by instruments mounted on this panel, which also carries the gas-flow control, pressure let-down, and isolating valves. The structure is of box form, with access to the inside for maintenance. Fig 4 shows the front view. On the other side is a test bench, with facilities for sampling the gas from a number of points in the stream and for making sulphur determinations, gas density measurements, etc.

The points in the main system at which pressures and temperatures are measured are shown in Fig 5. Details of the instruments are given below.

Pressure.—Six indicating gauges (Cambridge Instrument Co. Ltd.) connected to points in the low-pressure feed and residual gas lines; four indicating gauges (Budenberg Gauge Co. Ltd.) to points Nos. 7, 8, 9, and 15; three recording gauges (Drayton Regulator and Instrument Co. Ltd.) to Nos. 9, 21, and 23.

Temperature.—Two eighteen-line thermocouple switchboards (Elliott Bros (London) Ltd.); "Teloscale" indicator (Foster Instrument Co. Ltd.), to all temperature points; "Elliottronic" high-speed six-line recording potentiometer, with suppressed zero to give range 200° to 400° C (Elliott Bros) for reactor temperatures Nos. 12, 13, 14, 15, 16, and 17; "Elliottronic" recorder as above, range 0° to 600° C, for Nos. 3 (gas purifier), 20, 22, 23, 24, and 27.

Gas Flow.—Four low-pressure flowmeters (Rotameter Manufacturing Co. Ltd.) for water gas, hydrogen, and residual gas (two ranges); two high-pressure magnetically-linked-type Rotameters, for synthesis gas and recycle gas; one high-pressure recording flowmeter (Solway Flowrators Ltd.) for total inlet gas (synthesis + recycle gas).

Gas Composition.—Hydrogen recorder, conductivity cell type, range 45 to 75 per cent (Elliott Bros), for hydrogen in synthesis gas; three carbon monoxide percentage recorders (Infra-Red Development Co. Ltd.), range 25 to 50 per cent, for synthesis gas, 0 to 15 per cent for residual gas, and 0 to 400 p.p.m., with alarm bell at 160 p.p.m., for monitoring the plant building atmosphere.

Differential Pressure Panel

This is shown in the photograph, Fig 3, and a diagram of the gauges and purge gas connexions is given in Fig 6.

The system is arranged to measure the pressure difference, in inches of mercury, between the gas inlet and various levels in the reactor. The gauges are single-window type, with a 12-inch long groove in the back plate, painted white cellulose. The window is ¼-inch armourplate glass. A ball valve is fitted in the top connexion to minimize blow-out of mercury. Nos. 1 to 7 have a common mercury well. Purge gas flow for keeping the lines clear of catalyst is measured in 1 to 10-cu. ft/hr (S.T.P.) Rotameters.

A separate mercury U-gauge gives the pressure drop across the filters. The total pressure-drop across the reactor is indicated by a Budenberg "Microvar" gauge, range 0 to 50 p.s.i. (No. 11 in Fig. 6).

Gas Purifier

Organic sulphur compounds (amounting to 15 to 20 gr/100 cu. ft.) are removed from the combined gas stream, before compression, by passing through a bed of granules (approx ½-inch cubes) composed of 70 per cent iron oxide and 30 per cent sodium carbonate maintained at a temperature of 200° to 240° C. The

contact material is contained in a mild-steel cylinder (14 inches dia × 9 ft) with coned bottom, of capacity 9·6 cu. ft. (410 lb). It is heated by flue gas from an external furnace, and the synthesis gas is preheated in an annulus between the catalyst vessel and the flue-gas space. A catalyst charging hopper is fitted at the top and a discharging hopper at the bottom, each of capacity 0·6 cu. ft. (25 lb). "Treacle" valves are fitted between the catalyst vessel and the hoppers. The purified gas is cooled in an exchanger and a water-cooled coil. The oxide box in the outlet gas stream is designed to remove traces of hydrogen sulphide, which can, on occasion, be present in the gas after hot purification.

During the test runs the purifier was operated with a catalyst replacement rate of 25 lb/day.

Ancillary Equipment

Gas Compressor.—Four-stage, single-crank, hydrogen compressor, capacity 2500 cu. ft/hr to 400 atm (Reavell & Co. Ltd.).

Gas Recycle Pumps.—Three units, each consisting of two double-acting diaphragm pumps, with variable stroke mechanism; rated output of each unit, 240 cu. ft/hr measured at the working pressure (20 to 50 atm) against a pressure difference of 2 atm (E.C.D. Ltd., Tonbridge).

Heat Exchanger.—Cylindrical shell, $10\frac{3}{4}$ inches o.d. × 0·593 inch thick, with hemispherical ends, 11 ft 1 inch long overall, containing twelve 10 ft × 1 inch o.d. × $\frac{3}{4}$ inch i.d. tubes with 2 inches o.d. crimped gills at $\frac{3}{8}$-inch pitch, welded to tube plates;

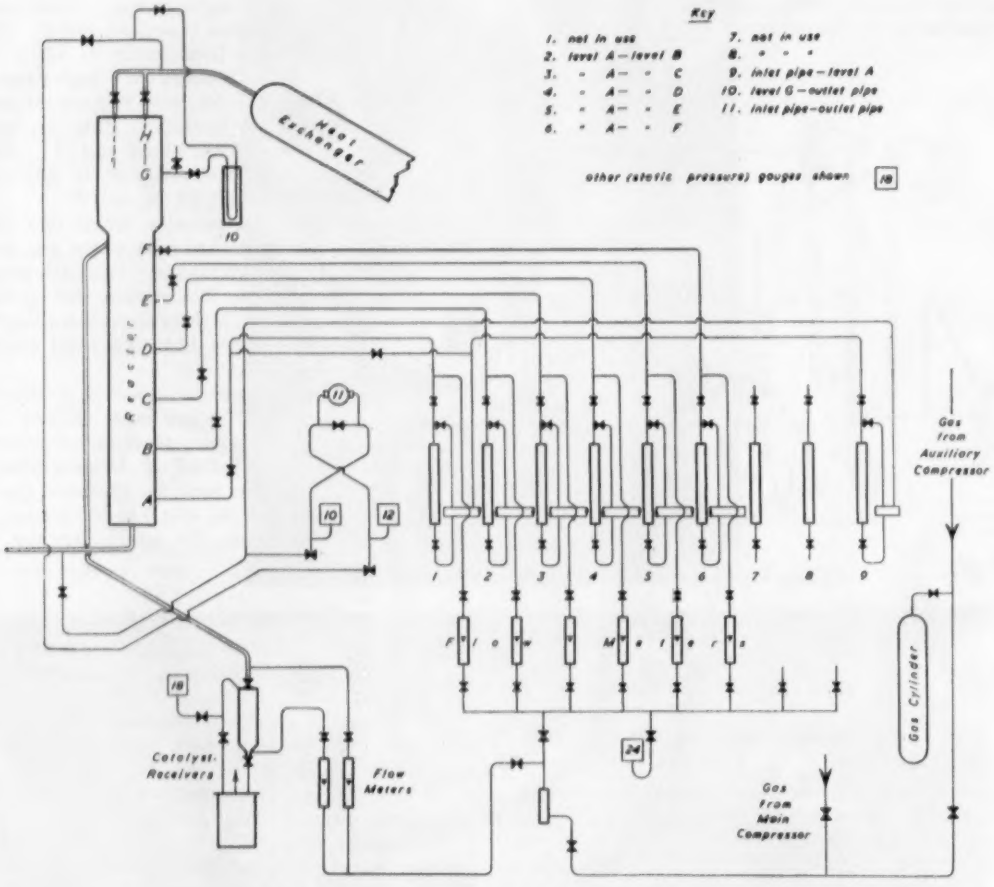


FIG 6

PILOT SCALE FLUIDIZED CATALYST HYDROCARBON SYNTHESIS PLANT DIAGRAM OF DIFFERENTIAL PRESSURE GAUGES AND PURGE GAS CONNEXIONS

tubes, tube plates, and ends 18/8/1 stainless steel; cylindrical shell and gills, mild steel; heat transfer surface (internal), 23·5 sq. ft. A 4-inch cleaning hole with cover was added at the lower end between Test Runs 1 and 2.

Gas Preheaters.—Three units, 5 ft, 5 ft, and 3 ft long, respectively, made of 3-inch i.d. × ¼-inch thick tube. Internal concentric tubes cause the gas to make three passes in the 5-ft lengths and two passes in the 3-ft length. They are fitted with electric heaters of 3, 3, and 2 kW respectively.

Final Condenser.—Cylindrical shell, 10¾ inches o.d. × 0·593 inches thick, with hemispherical ends, 11 ft 1 inch overall, containing thirty 10 ft × 1-inch o.d. × ¾-inch i.d. tubes welded to tube plates; tubes, tube plates, and ends, 18/8/1 stainless steel, cylindrical shell, mild steel; heat transfer surface (internal), 59 sq. ft. A 4-inch cleaning hole with cover was added between Test Runs 1 and 2.

High Pressure Gas–Liquid Separator.—Cylinder, 7⅞ inches i.d. × ⅜ inch thick × 3 ft 9 inch long, with flanged end covers, in mild steel. It is fitted with a double-glass liquid level gauge.

Water-heater.—Mild steel cylinder, 10¾ inches o.d. × 0·593 inch thick × 2 ft 4 inches long with hemispherical top and flanged bottom cover fitted with two 3-kW immersion heaters and two 2-kW external heaters on the cylindrical portion.

Steam Chest.—Cylindrical shell, 10¾ inches o.d. × 0·593 inch thick, with hemispherical ends, 6 ft 4 inches long overall, containing twelve 3 ft 0 inch × ¾-inch o.d. × ½-inch i.d. tubes, with 1¾-inch o.d. crimped gills at ⅜-inch pitch between tube plates in the upper portion, all in mild steel; heat transfer surface (internal), 4·7 sq. ft. It is fitted with a water level gauge.

Catalyst Overflow System.—This is shown diagrammatically in Fig 6. The high-pressure receiver is 6 inches i.d. × 23 inches long, and is connected to the reactor by ¾-inch bore pipe. Klinger ¾-inch sleeve-packed cocks are fitted at the top and bottom. The lower valve discharges the catalyst, via a sight glass, to a low-pressure receiver, which is vented, through a sintered plate filter, to the residual gas line. A pressure release line is connected between the two vessels, and fluidizing gas is supplied at the lower end of the ¾-inch pipe and to the base of the higher-pressure receiver.

Evaporator for Process Steam.—A 16-ft length of ⅜-inch bore 18/8/1 steel tubing, with five heater windings totalling 19·5 kW. It is fed with measured distilled water by a piston pump.

Piping, Valves, etc.—The piping used for the main high pressure gas and cooling system lines is either ½-inch or ¾-inch bore, except for a length of 1½-inch bore between the high-pressure separator and the recycle pumps. Connexions to gauges, sampling points, etc., are $\frac{3}{16}$-inch bore. Ermeto couplings are used generally throughout. Stainless 18/8/1 steel is used for the short connexion between the heat exchanger and the final condenser, otherwise all piping is mild steel. Most of the valves are either Ermeto bonnet type, Ermeto gas control (needle), or Klinger AB10.

EXPERIMENTAL DETAILS

The Catalyst

An alkali impregnated mill-scale catalyst of the type used in experiments in the smaller reactors [3–5] was employed. Mill scale, obtained from the rolling mills of the Ford Motor Co. at Dagenham, Essex, was ground in a ball mill and screened to pass a No. 72 and remain on a No. 170 BS Test Sieve. The graded scale was impregnated with potassium carbonate by mixing with a dilute aqueous solution and drying in a steam-heated dough-mixer. The material was then loaded into fireclay saggers and heated for 2 hours at 1000° C in a gas-fired muffle furnace. The composition of the finished catalyst, in percentages by weight, was as follows: Fe_2O_3, 96·0; K_2O, 1·8; SiO_2, 0·7; Cr_2O_3, 0·6; V_2O_5, 0·6; Mn_3O_4, 0·2; Al_2O_3, 0·1.

Computation of Results

For the assessment of performance, the runs were divided into test periods each of approximately 24 hours' duration. During each such period, the corrected volumes of feed gas and exit gas were obtained from readings of the wet meters and the weight of condensed oil and water collected in the product receiving tanks was determined. The amount of light hydrocarbons (C_5–C_7) was determined by passing a portion of the exit-gas stream through traps immersed in dry ice (carbon dioxide) and weighing the contents of the traps after stabilization at room temperature. Average samples of feed and exit gas were taken in 5-litre aspirator bottles and analysed in a Bone-and-Wheeler type apparatus. From these analyses and the observed volume contraction during synthesis, the percentage conversions of carbon monoxide and hydrogen and the proportions of converted carbon monoxide appearing as carbon dioxide, methane, and higher hydrocarbons were computed.

In selected periods a more complete product distribution was obtained by analysing a sample of exit gas in a "Robot" Podbielniak apparatus. Two methods were used to obtain samples for these analyses. In one, a portion of the exit-gas stream was freed from carbon dioxide and the hydrocarbons were liquefied by passage through a series of traps immersed in liquid nitrogen. The condensate was then evaporated into a 20-litre aspirator bottle using saturated calcium chloride solution as confining liquid. In the other (the preferred procedure) a 1 to 1½-cu. ft. sample of exit gas (after removal of CO_2) was collected in a glass carboy and the gas liquefied into the Podbielniak still.

Yields of acidic and neutral organic compounds in solution in the reaction water were obtained from the weight of reaction water and the acid content (determined by potentiometric titration) and a determination of alcohols and other neutral compounds based on

ether-extraction of a sample of the water after saturation with potassium carbonate.

A daily sample of total non-aqueous liquid product was obtained by blending the light spirit obtained by refrigeration of the exit gas with the appropriate weight of condensed oil from the receiving tank. The boiling range of the total product was obtained from an Engler-type distillation, and an analysis for active groups was carried out on the fraction distilling 80° to 150° C. In certain cases the calorific value and the total oxygen content (by the Schutze–Unterzaucher method [6]) were determined on samples of total non-aqueous product.

Samples of catalyst from the reactor were exhaustively extracted with benzene and analysed for total iron and total carbon. From the iron and carbon contents, the proportions of iron oxide, iron carbide, and uncombined carbon were computed on the assumption that, after use in the synthesis, the iron is present only as Fe_3O_4 and Fe_2C. This has been established by X-ray and thermo-magnetic studies of iron catalysts carried out by German and American workers and confirmed in a number of cases at the Fuel Research Station.

PRELIMINARY TRIAL RUN

Although in previous work on a smaller scale the mill-scale catalyst was reduced in pure hydrogen at 450° C for a period of 24 hours, it had been established that this catalyst could be completely reduced at much lower temperatures (*e.g.* in a period of 120 hours at 290° C). For the preliminary trial run, therefore, an attempt was made to reduce the catalyst *in situ*, at the maximum temperature attainable by using the electric heaters in the water-heater of the reactor cooling system and the feed gas electric pre-heaters.

The reactor was charged with 138 lb of catalyst and the reduction carried out in hydrogen prepared by the steam–iron process (containing 0·6 per cent CO_2 and 2·5 per cent CO) passed at 1500 cu. ft/hr (S.T.P.) and 5 atm gauge pressure. The compressor separator was maintained at 70 atm pressure to dehydrate the hydrogen. The reduction was continued for a period of 122 hours with a catalyst temperature of 283° to 292° C. The plant pressure was then increased to 20 atm with hydrogen passing at 1500 cu. ft/hr and the recycle pumps were set to give a total rate of *ca* 5000 cu. ft/hr (S.T.P.). With the catalyst temperature at 285° C, the flow of water-gas was started at 250 cu. ft/hr, and an immediate increase in catalyst temperature resulted. During the next 12 hours the water-gas rate was increased gradually to 1200 cu. ft/hr and the hydrogen rate was reduced to 800 cu. ft/hr. The automatic cooling-water pressure regulator was brought into action and the catalyst temperature levelled out smoothly at 310° to 315° C. At this stage the volume contraction and readings of the infra-red gas analyser recorders indicated a carbon monoxide conversion of 85 per cent. After fairly steady running for 48 hours, mechanical failure of the recycle pumps necessitated a shut-down. After making temporary repairs to the pumps, synthesis conditions were re-established and maintained for a period of 72 hours, with interruptions to repair leaks on the water-heater of the cooling system. At 20 atm pressure, catalyst temperature 310° to 315° C, feed gas rate 2100 cu. ft/hr, the carbon monoxide conversion was only 67 per cent, and due to persistent failure of one or more of the recycle pumps the recycle ratio could not be maintained above 0·7 : 1. The run was therefore terminated.

On inspection of the interior of the reactor it was found that the layer of steel balls had become cemented together with catalyst and there was a loosely-compacted mass of catalyst for a distance of 12 to 18 inches up one side of the reactor between the wall and the cooling tubes. It appeared that an inadequate depth of balls combined with a low linear velocity (due to recycle pump failures) had led to non-uniform gas distribution and incomplete fluidization of the bed.

During the first 24 hours under synthesis conditions, the pressure-difference panel was put out of action by the condensation of liquid hydrocarbons in the lines and rotameters from the recycle gas used as purge. A supply of compressed hydrogen from an auxiliary compressor was therefore installed to provide purge gas in the next run. The recycle pumps were returned to the makers for repairs and modifications, and the inlet and outlet manifolds were increased in diameter to reduce excessive surging in the recycle-gas stream.

Analysis of a sample of catalyst taken from the reactor before starting synthesis indicated that reduction had been only 40 per cent complete. Tests in the small reactors showed that this was due to the presence of CO and CO_2 in the hydrogen employed. For the next run it was decided to use electrolytic hydrogen and to carry out reduction at 400° to 450° C. To permit this, electric heating windings were fitted to the reactor walls. A number of other minor modifications were also made as a result of experience during the trial run.

TEST RUN 1

For the first test run it was decided to employ the sintered-metal gas distributer already described. It was tested by fitting the bottom cover of the reactor, incorporating the distribution plate, to a 5-ft length of 8-inch dia Pyrex glass pipeline and fluidizing a 3 to 4-ft bed of unreduced catalyst in air. The reactor cooling-tube system was suspended in the bed. At linear velocities for which the pilot plant was designed (0·4 to 0·6 ft/sec) uniform gas distribution and smooth fluidization were obtained.

The catalyst charge for this run comprised 95 lb of catalyst recovered from the trial run (containing 2·5 per cent of benzene-soluble matter and 6·9 per cent total carbon) and 80 lb of fresh, mill-scale catalyst.

For the reduction, 4000 cu. ft. of electrolytic hydro-

gen was discharged from cylinders into a thoroughly-purged gas holder and circulated through the reactor via the main compressor. The water was discharged from the reactor cooling system, the plant pressure was increased to 5 atm with hydrogen circulating at 1600 cu. ft/hr (S.T.P.) and the catalyst temperature raised to 420° C by a combination of direct heat on the reactor wall and sensible heat of the preheated temperature raised to 250° C, with hydrogen flowing at 1500 to 1800 cu. ft/hr (s.t.p.). Flow of water-gas was started at 500 cu. ft/hr, and the rate was increased during the ensuing 12 hours (with reduction in the hydrogen flow to maintain 2000 cu. ft/hr total fresh feed) to give a mixture of H_2–CO = 2·5. The catalyst temperature levelled out at 320° C, and the CO recorders indicated a gas conversion of the order

TABLE I

Operating Conditions Used and Conversions and Yields Obtained in Test Run 1

Test period	2	4	6	8	10	12	15
Synthesis gas :							
CO + H_2, %	95·8	96·9	94·7	94·7	96·9	95·9	94·9
H_2–CO ratio,	2·25	2·17	1·95	1·93	1·87	1·93	1·92
Total sulphur, gr/100 cu. ft.	0·05	0·07	0·06	—	—	0·06	—
Input—							
cu. ft/hr (S.T.P.)	2126	2157	2160	2080	2146	2209	2057
cu. ft/cu. ft. fluid catalyst/hr	1690	1710	1480	1320	1310	1345	1250
cu. ft/lb iron/hr	18·0	—	—	—	—	—	19·7
Recycle ratio, cu. ft. tail gas/cu. ft. fresh feed	1·5	1·5	1·5	2·6	2·4	2·2	2·5
Linear gas velocity in reactor, ft/sec	0·59	0·60	0·61	0·59	0·59	0·56	0·57
Operating pressure :							
Reactor inlet, atm, gauge	21·0	21·2	21·0	29·5	29·2	29·5	29·6
Reactor outlet, atm, gauge	20·0	20·2	20·0	28·5	28·2	28·5	28·7
Steam pressure in cooling system, atm, gauge	71·0	71·0	71·0	67·5	67·5	67·0	70·5
Fluid catalyst volume, cu. ft.	1·26	1·26	1·46	1·58	1·64	1·65	1·65
Temperature :							
Reactor inlet, ° C	327	323	299	302	300	305	307
Catalyst—							
Level A, ° C	326	328	327	327	329	327	325
Level B, ° C	323	326	325	325	326	324	322
Level C, ° C	321	323	322	322	324	322	322
Level D, ° C	314	316	320	319	321	319	319
Reactor cooling water, ° C	289	289	289	285	285	285	288
Gas volume contraction, %	65·5	66·3	64·9	76·4	77·3	72·5	69·6
CO in fresh feed converted, %	93·7	92·8	93·2	96·0	97·1	94·9	90·8
(CO + H_2) in fresh feed converted, %	80·6	81·5	82·8	91·5	92·8	89·1	83·7
Usage ratio, moles H_2–mol CO	1·70	1·69	1·53	1·73	1·68	1·63	1·63
Distribution of converted CO :							
To CO_2 %	9·5	10·9	15·4	8·7	9·1	6·5	7·8
To CH_4, %	12·2	11·6	11·7	10·6	10·0	13·9	11·1
To higher hydrocarbons, %	78·3	77·5	73·9	81·7	80·9	79·6	81·1
Observed yields of products :							
Condensed oil, g/m³ (S.T.P.) fresh feed	42·4	39·8	39·1	55·6	49·8	35·1	31·4
Light spirit in tail gas, g/m³ (S.T.P.) fresh feed	8·0	10·7	16·2	14·5	15·3	10·2	11·1
Condensed water, g/m³ (S.T.P.) fresh feed	176·1	168·4	166·3	211·3	207·5	208·1	191·2
Acids * in water, g/m³ (S.T.P.) fresh feed	4·2	—	5·0	—	7·2	7·1	—
Neutral organic matter in water, g/m³ (S.T.P.) fresh feed	14·6	—	14·6	—	19·1	22·4	—

* Calc as acetic acid.
Test periods of 24 hours duration, except Period 15, which was only 6 hours.
Operating pressure increased from 20 to 30 atm during Period 7.
Synthesis was interrupted between Periods 11 and 12 and between 14 and 15.
Up to Period 5 the catalyst level was between C and D; from Period 6 onwards it was between D and E.
The locations of Levels A, B, C, etc., are shown in Figs 2 and 6.

hydrogen. The reduction was continued for a period of 24 hours from the point when the catalyst temperature reached 420° C. During this period the catalyst temperature rose slowly, and during the final 6 hours was 440° to 445° C. The fall in pressure-drop across the catalyst bed indicated that reduction was complete; this was confirmed, later, by analysis of a sample of catalyst removed from the reactor.

The reactor cooling system was re-charged with distilled water, the plant pressure was increased to 20 atm in steam–iron hydrogen, and the catalyst of 90 per cent. The first test period was started 24 hours after the flow of water-gas was started. After six days' steady running at 20 atm with the H_2–CO ratio of the feed gas reduced in stages to 1·95 : 1, and with some increase in catalyst temperature, the operating pressure was increased to 30 atm. After a further five days' satisfactory performance, a gas leak and fire at the triple-flange joint at the outlet of the reactor cooling-system caused a shut-down. The leak was repaired, and synthesis conditions were re-established and maintained for 52 hours. A rapidly

rising pressure-difference across the product-outlet system then necessitated a further shut-down. The build-up of pressure in the outlet system was traced to a blockage in the product line between the heat exchanger and final condenser, consisting of basic ferric acetate. Early in the run evidence had been obtained that catalyst dust was present in the product stream, and after the shut-down a quantity of catalyst was found to have accumulated in the lower end of the heat exchanger and the upper end of the condenser. The iron salt appears to have been derived from the reaction of acetic acid in the reaction water with the iron in the catalyst dust. After clearing the blockage, synthesis conditions were re-established but some 6 hours later the blockage reformed at the same point and the run was terminated.

After cooling down, the catalyst was discharged through the sample cock in the bottom cover. Top and bottom covers were then removed and the reactor was found to be clean. Some catalyst dust was found adhering to the filters, and a small quantity was recovered from the overflow receiver. An estimate of the total amount of catalyst which had left the system in the liquid products was obtained from

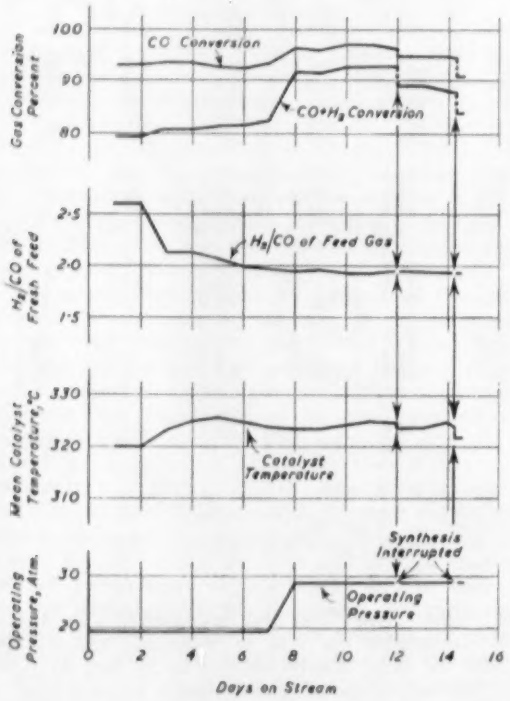


FIG 7

VARIATION IN OPERATING CONDITIONS AND GAS CONVERSION WITH TIME—TEST RUN 1

determination of the amount present in 2-litre samples withdrawn from the oil–water separator from time-to-time during the run.

The operating conditions, gas conversions, product yields, and other data for seven of the fifteen test periods are shown in Table I, and variations in

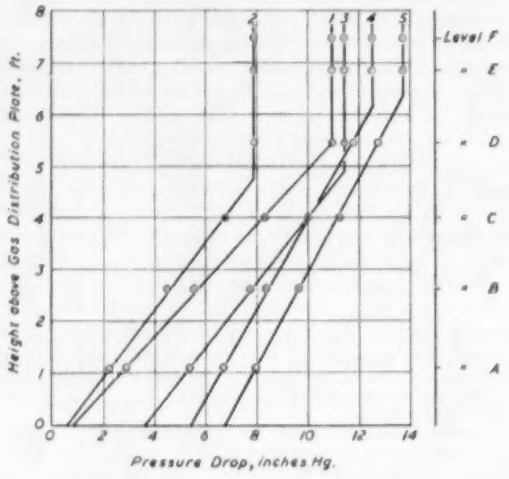


FIG 8

PRESSURE DROP AT VARIOUS LEVELS IN REACTOR—TEST RUN 1

1. In H_2, before reduction.
2. In H_2, after reduction.
3. Synthesis, Period 4.
4. Synthesis, Period 8.
5. Synthesis, Period 15.

operating variables and conversion of reactants with time are shown graphically in Fig 7. The best performance was obtained during the five days following the increase in pressure from 20 to 30 atm when the CO conversion was 96 to 97 per cent and the (CO + H_2) conversion 92 to 93 per cent. After each shut-down the fall in gas conversion was steeper than would have been expected as a result of normal deterioration of the catalyst with time. Yields of gaseous hydrocarbons and oxygen-containing water-soluble products were higher than had been anticipated, and the latter appeared to increase with time of operation in addition to the expected increase with increase in working pressure. It will be noticed that the catalyst temperature decreased by 5° to 8° C from the bottom to the top of the catalyst bed.

Data derived from readings of the pressure-difference gauges are shown graphically in Figs 8 and 9. The linearity of the plots in Fig 8 indicates that the fluidized bed was of uniform density throughout its depth. The intercept on the PD axis gives the pressure drop across the gas distribution plate, which increased with increase in the density of the gas. The height of the fluid bed is indicated by the point of intersection of the vertical and sloping portions of the PD plots. It will be noted from Fig 9 that little expansion of the catalyst took place during the first five days, but that thereafter a fairly rapid rise

(apparently following reduction in the H_2–CO of the feed gas to *ca* 2·0) took place, the bed height finally becoming constant at $6\frac{1}{2}$ ft. The density of the bed fell during the entire run from 110 to 80 lb/cu. ft.

The amount and composition of the solids recovered from the system are shown in Table II. It can be calculated from these data that if all the solids found had remained in the bed the final depth of bed would have been about $7\frac{1}{2}$ ft, *i.e.* just above the catalyst overflow point. The total carbon formed corresponds to an average rate of 2·45 g/m³ CO converted. This is of the order expected from the partial pressure of hydrogen in the reactor, which averaged 195 p.s.i. during the run (see Fig 7 of reference 4). The analysis of the main bulk of catalyst found in the reactor showed that little oxidation of the iron had occurred and reveals no cause for the final fall in activity. The sulphur content of the discharged catalyst was 0·08 per cent. The analyses indicate that of the total carbon formed only 60 per cent was uncombined carbon.

Properties of the liquid products obtained at various stages of the run are shown in Table III. The boiling range of the product did not vary in any consistent manner, but the oxygen content of the 80° to 150° fraction (approximately 30 per cent of the total) increased with time, a trend which is consistent with the increased yield of water-soluble oxygen-containing products. The oxygen content of an average sample of the total liquid product produced during the run was 8·5 per cent by weight.

From Podbielniak analyses of the gaseous hydrocarbons and data given in Tables I and III, the more complete product distributions and yields shown in

TABLE II

Weight and Composition of Catalyst Charged and Recovered, Test Run 1

Description	Weight,* lb	Bulk density, lb/cu. ft.	Total iron		Total carbon		Main constituents, % by w		
			%	lb	%	lb	Fe_3O_4	Fe_3C	"Free C"
Unreduced catalyst as charged	175	150	—	—	—	—	—	—	—
Reduced catalyst	128	125	92·2	118·0	1·0	1·28	—	—	—
1. Catalyst found in reactor at the end of the run	144	112	72·6	104·5	20·1	28·9	3·3	77·7	12·6
2. Catalyst found in heat exchanger and condenser	17·4	63	43·9	7·6	27·7	4·9	—	—	—
3. Catalyst found in overflow receiver	6·6	115	70·3	4·6	19·0	1·3	16·8	64·3	12·8
4. Catalyst in liquid products	2·5	—	34·3	0·9	47·2	1·2	—	—	—
5. Catalyst adhering to filters	1·5	58	67·1	1·0	21·8	0·3	19·3	58·8	16·1
Total recovered	172·0			118·6		36·6			

* All weights refer to material insoluble in benzene. Samples 1, 3, and 5 contained no benzene-soluble material.

TABLE III

Properties of Liquid Products (Condensed oil + Spirit), Test Run 1

Test period	2	4	6	8	10	12	15
Synthesis conditions:							
H_2–CO of feed gas	2·25	2·17	1·95	1·93	1·87	1·93	1·92
Operating pressure, atm, gauge	20	20	20	29	29	29	29
Mean catalyst temperature, °C	323	326	324	323	325	324	322
Boiling range, % by w:							
Up to 200° C	80·5	83·1	87·4	86·4	86·9	89·3	81·6
200° to 300° C	12·8	12·3	10·0	9·2	9·6	8·4	11·5
Residue at 300° C	6·7	4·6	2·6	4·4	3·5	2·3	6·9
Analysis of 80°–150° *fraction* *:							
Olefins, % by w	59·1	56·6	52·4	53·0	51·7	44·5	35·0
OH number, mg KOH/g	74·5	84·4	107·4	128·0	127·7	149·3	216·8
Acid, mg KOH/g	22·3	27·6	29·7	34·0	36·4	40·3	42·8
Ester, mg KOH/g	20·1	21·5	25·4	30·0	68·0	40·5	79·4
Total oxygen, %	4·5	5·2	6·2	7·3	9·6	8·9	13·2

* Approx 30% of the total.

From Podbielniak analyses of the gaseous hydrocarbons and data given in Tables I and III, the more complete product distributions and yields shown in

Table IV, for two test periods, were obtained. Test Period 10 is representative of the peak performance achieved. It will be noted that of the total product obtained (183 g/m³ feed gas) methane and C_2 hydrocarbons, which are not regarded as recoverable

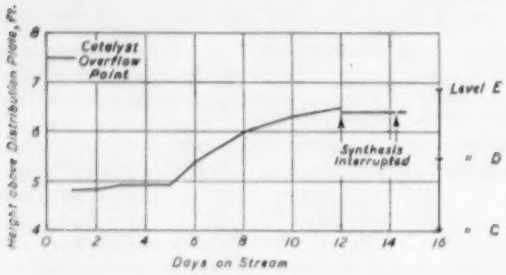


FIG 9

CATALYST BED EXPANSION WITH TIME—TEST RUN 1

products, amounted to over 25 per cent and that, of the recoverable products (C_3 and higher), nearly 20 per cent consisted of water-soluble chemicals. Propylene and butylenes, which together amounted to over 20 per cent of the total product, are, clearly, major products of this type of operation.

TABLE IV

Product Compositions and Yields for Selected Periods of Test Run 1

Test period	6		10	
Synthesis conditions:				
H_2–CO of feed gas	1·95		1·87	
Space velocity, vol/vol fluid catalyst/hr	1480		1310	
Mean catalyst temperature, ° C	324		325	
Operating pressure, atm, gauge	20·5		28·7	
	g/m³ feed gas	per cent of total	g/m³ feed gas	per cent of total
Total products:				
Methane	25·2	15·8	24·5	13·4
Ethylene	11·6	7·3	9·5	5·2
Ethane	8·0	5·0	12·8	7·0
Propylene	21·8	13·6	22·0	12·0
Propane	2·2	1·4	4·4	2·4
Butylenes]	13·0	8·1	16·0	8·8
Butanes	3·3	2·0	2·4	1·3
C_5 and higher	55·3	34·6	65·1	35·6
Acids in reaction water	5·0	3·1	7·2	3·9
Neutral organic matter in reaction water	14·6	9·1	19·1	10·4
Total	160·0	100·0	183·0	100·0
Recoverable, "useful" products:				
Propane and butanes	5·5	4·8	6·8	5·0
Propylene and butylenes	34·8	30·2	38·0	27·9
Crude naphtha, B.P. 200° C	48·3	41·9	56·5	41·5
Middle oil, 200°–300° C	5·5	4·8	6·3	4·6
Heavy oil and wax	1·5	1·3	2·3	1·7
Acids and neutral chemicals	19·6	17·0	26·3	19·3
Total	115·2	100·0	136·2	100·0
Output of "useful" products:				
lb/day	370		436	
lb/cu. ft. fluid catalyst/day	254		265	
gal/day	53		62	

TEST RUN 2

Before the next run was carried out, the system for injecting pre-heated steam into the feed gas (shown in Fig 5) was installed. In the 1-inch diameter unit promising results had been obtained [7] in runs with hydrogen-deficient gas such as blue water-gas and producer gas with the addition of steam to minimize carbon formation which is otherwise unacceptably high when using such gases. Before investigating this on the pilot-plant scale, however, it was considered desirable, in view of some of the unsatisfactory features of Run 1, to start the second run under "conventional" operating conditions.

For this run, residual gas stripped of liquid hydrocarbon vapour by passage through active carbon was used as purge gas for the pressure-difference gauge lines, catalyst overflow, and filter blow-back systems. The filter system was altered, as indicated in the description of the plant, before starting the run.

The reactor was charged with 212 lb of freshly-prepared mill-scale catalyst and the reduction carried out in electrolytic hydrogen under closely similar conditions to those used in Run 1. The fall in PD across the bed (and subsequent analysis of a sample of the catalyst) showed that reduction had proceeded to completion.

Synthesis was initiated, as in the previous run, by admitting water-gas at a gradually increasing rate with the bed fluidized in hydrogen at 20 atm pressure and 260° C. Synthesis at a mean catalyst temperature of 322°, with H_2–CO of feed gas *ca* 2·0 was established 12 hours later and the first test period started. Operation, substantially under these contions, was continued for eight days, when it became necessary to shut-down to repair a leak of steam at the outlet of the reactor cooling system. During Period 5 the temperature of the products leaving the final cooler had risen from 29° to 40° C, and after the shut down this was found to be due to wax solidifying in and blocking some of the condenser tubes. The heated condenser by-pass line, to allow molten wax to flow direct from the exit of the heat exchanger to the product collecting system, was therefore fitted before re-starting synthesis. For the next seven days (Test Periods 9 to 15) the plant ran smoothly, apart from inefficient operation of the recycle pumps. The low recycle ratio led to a low (CO + H_2) conversion, and although this improved when the recycle ratio was finally maintained for five days at 1·7 : 1, the decision to increase the operating pressure to 25 atm after Test Period 15, and to make a further increase to 30 atm after Period 18, was made primarily in order to increase this conversion. The CO conversion had remained above 95 per cent from the start of the run. The (CO + H_2) conversion did not, however, exceed 90 per cent, and, in fact, after the second increase in pressure it started to fall. During this part of the run (Periods 19 to 21) the steam joint at the reactor cooling system outlet began to leak again, and at the conclusion of Period 22 (which was of only 5 hours' duration) it became necessary to shut-down.

At the end of Period 12, 10 lb of catalyst had been recovered from the overflow receiver. This was assumed to be a gradual accumulation during the

proceeding 12½ days operation, but at the end of Period 16 a further 15 lb was found in the receiver, although the bed level indicated by the PD gauge readings was some 18 inches below the overflow point. The inlet to the catalyst receiver was, therefore, closed for the next 24 hours and then opened for a period of ½ hour. Only a small amount of catalyst was then found in the receiver, indicating that the true bed level was, as the PD readings indicated, made no difference to the temperature, but a succession of sharp blows on the top of the reactor caused the temperature to fall to the normal value. It was assumed, therefore, that a loose accumulation of fine catalyst, on which synthesis was taking place, had formed around the filter banks.

The recycle pumps were overhauled and new diaphragms fitted. The PD panel was overhauled and some partly blocked lines were cleared. Blockage of these lines had been suspected from the erratic readings obtained during the last few days before the shut-down.

After repairs had been completed, the bulk of the recovered catalyst (including some recovered from the overflow receiver) was replaced in the reactor and (in view of the evidence of falling activity before the shut-down) was reduced in hydrogen at 400° to 430° C and 5 atm pressure for 24 hours. Synthesis conditions at 20 atm were established with a feed gas of H_2–CO approx 4, and a 24-hour Test Period (No. 23) carried out with this gas to provide an indication of the performance which might be obtained if re-formed coke-oven gas was used as feed. The maximum catalyst temperature attainable with this gas at a steam pressure of 70 atm was 305° C.

A change was then made to normal feed gas

TABLE V

Operating Conditions, Conversions, and Yields obtained in Test Run 2

Test period	2	4	6	8	10	12	14	16	18	20	23	25	27	28	29	30	31
Synthesis gas (H_2 : CO)	2·08	1·99	1·99	2·31	2·04	2·00	1·90	1·96	1·96	1·97	3·78	1·97	2·00	1·03	1·17	1·16	1·84
CO + H_2, %	97·1	97·9	96·1	97·4	96·0	97·2	96·7	95·2	96·1	97·7	98·0	96·3	95·4	93·4	92·1	92·1	93·9
Total sulphur, gr/100 cu. ft.	0·06	0·01	0·02	—	0·02	—	0·04	—	0·06	—	—	0·08	—	—	0·07	—	—
Input, cu. ft/hr	2060	2050	2050	2103	2104	2096	2104	2086	2100	2100	1880	2137	2175	1660	1632	1680	2100
Input, cu. ft/cu. ft. fluid catalyst/hr	1380	1330	1310	1330	1330	1330	1330	1330	1330	1330	1170	1250	1210	925	900	925	1150
Input, cu. ft/lb reduced iron/hr	15	—	—	—	16	—	17	—	18	19	17	19	19	15	15	15	19
Process steam, lb/hr	—	—	—	—	—	—	—	—	—	—	—	—	—	22·8	22·4	25·8	—
Process steam, g/m³ feed gas	—	—	—	—	—	—	—	—	—	—	—	—	—	226	222	246	—
Recycle ratio, cu. ft tail gas/cu. ft. fresh gas	1·2	1·2	1·2	1·7	1·2	1·7	1·6	2·2	2·4	2·2	1·0	1·25	2·2	2·3	2·5	2·4	2·4
Linear gas velocity in reactor, ft/sec	0·52	0·52	0·52	0·65	0·53	0·60	0·63	0·61	0·66	0·52	0·43	0·55	0·53	0·51	0·54	0·52	0·55
Operating pressure :																	
Reactor inlet, atm, gauge	21·9	22·0	22·2	22·2	21·8	21·8	21·9	26·0	26·4	29·5	20·5	20·8	30·0	29·9	29·9	29·9	30·0
Reactor outlet, atm, gauge	20·3	20·2	20·6	20·2	20·2	21·1	21·0	25·2	25·1	28·3	20·0	20·1	29·1	29·0	29·0	29·0	29·1
Steam pressure in cooling system, atm, gauge	68·0	68·2	67·7	68·1	68·0	71·5	71·7	71·7	72·5	73·3	71·9	76·3	76·2	75·5	78·8	85·7	85·5
Fluid catalyst volume, cu. ft.	1·49	1·54	1·56	1·58	1·58	1·58	1·58	1·58	1·58	1·58	1·61	1·71	1·80	1·80	1·80	1·82	1·86
Temperatures :																	
Reactor inlet, ° C	279	279	279	287	308	306	306	303	303	309	287	287	281	299	306	297	306
Catalyst—																	
†Level A, ° C	329	329	327	328	328	329	329	330	330	330	308	323	326	320	321	324	320
Level B, ° C	324	323	321	324	325	324	326	327	327	327	304	320	322	317	317	320	317
Level C, ° C	320	320	319	322	320	322	323	325	325	325	303	318	321	316	316	320	316
Level D, ° C	317	317	316	320	318	320	321	322	323	323	302	317	320	314	314	318	315
Reactor cooling water, ° C	286	286	285	286	286	289	289	289	290	291	289	293	293	293	293	302	302
Gas volume contraction, %	74·4	76·3	73·3	77·6	74·4	77·6	77·9	77·6	76·9	73·5	44·5	60·2	69·9	44·3	46·6	41·7	55·2
CO in fresh feed converted, %	97·5	98·2	97·5	97·3	97·2	97·4	96·9	96·6	96·9	95·7	89·8	91·7	94·4	90·3	89·5	87·1	81·8
(CO + H_2) in fresh feed converted, %	86·9	88·3	87·3	90·0	86·5	89·2	90·2	90·1	89·3	86·7	54·5	73·5	84·4	71·5	72·3	66·9	70·1
Usage ratio, mols H_2/mol CO	1·75	1·69	1·65	2·07	1·71	1·75	1·70	1·76	1·72	1·69	1·90	1·39	1·68	0·61	0·75	0·66	1·43
Distribution of converted CO :																	
To CO_2, %	9·2	9·9	8·6	8·4	8·5	6·4	8·4	8·0	7·0	7·9	12·5	15·1	13·0	34·5	40·3	43·4	16·3
To CH_4, %	10·5	8·7	10·1	7·8	8·6	8·8	7·0	7·8	7·2	9·8	15·7	8·8	10·1	6·8	6·6	5·9	12·9
To higher hydrocarbons, %	80·3	81·4	81·3	83·8	82·9	84·8	84·6	84·2	85·8	82·3	71·8	76·1	76·9	58·7	53·1	51·7	70·9
Observed yields of products :																	
Condensed oil, gm³ (S.T.P.) fresh feed	74·9	75·2	66·7	66·6	57·1	54·0	69·4	84·6	53·9	57·1	17·7	56·5	45·9	37·0	38·9	30·6	34·7
Light spirit, g/m³ (S.T.P.) fresh feed	9·9	11·6	15·8	12·8	21·2	29·0	20·0	23·7	18·2	18·0	5·4	12·1	12·8	13·2	11·8	18·3	22·5
Condensed water, g/m³ (S.T.P.) fresh feed	208·6	214·0	214·0	227·5	226·5	218·4	212·4	217·6	215·5	231·6	123·2	212·9	181·0	267·0	252·0	272·0	146·0
Acids* in water, g/m³ (S.T.P.) fresh feed	5·5	6·0	6·5	6·8	6·7	7·0	6·8	7·0	8·2	8·1	3·5	—	6·7	12·0	12·9	14·2	—
Neutral organic matter in water, g/m³ (S.T.P.) fresh feed	14·1	14·7	17·5	17·0	20·2	21·4	19·2	19·5	20·2	25·7	12·3	—	20·3	21·9	17·1	15·3	—

* Calc as acetic acid.

† The locations of Levels A, B, C, etc., are shown in Figs 2 and 6.

below the overflow point and that the catalyst previously found in the receiver had originated from " plumes " of catalyst thrown up from the bed.

To make an effective repair of the steam joint it was necessary to discharge the catalyst and remove top and bottom reactor covers. The catalyst was discharged as a free-flowing powder, and the interior of the reactor was found to be clean and free from catalyst accumulation. The filters were coated with a thin layer of fine catalyst dust with some bridging between neighbouring filters. There had been no increase in PD across the filters during the run; blowing back, having made no difference to the PD, had not been practised since the early periods of the run. During Period 15, however, the temperature of the gas leaving the reactor was found to be higher than that at the top of the reactor (Level H). Blowing back the filters

(H_2–CO = 2) and the operating pressure maintained at 20 atm for Periods 24 and 25 with the steam pressure increased to 76 atm to maintain a catalyst temperature of 320° C. Gas conversion was lower than before the shut-down, and the operating pressure was increased to 30 atm for Periods 26 and 27. This restored the conversion to about the level obtained in Periods 20 and 21, suggesting that the hydrogen treatment had produced little activating effect on the catalyst. Before Period 28 the supply of hydrogen was cut off, leaving the water-gas rate unchanged, and distilled water was pumped into the process steam system at about 2 gal/hr. After two periods, the supply of steam was increased slightly, and the pressure in the reactor cooling system was increased to 85 atm to offset the fall in gas preheat and catalyst temperatures. The steam injected during these periods brought the hourly rate of total feed to the same level (2050 to 2150 cu. ft.) as that during normal synthesis. The steam constituted 21 to 23 per cent of the total feed. In Period 31 a return to normal operation with water-gas and hydrogen was made in order to assess catalyst activity. The performance was appreciably worse than before the steam injection, and after Period 32 a further treatment with hydrogen at 400° to 440° C for 24 hours was carried out. The activity of the catalyst was then so low that no rise in catalyst temperature resulted when water-gas preheated to 300° C was supplied to the reactor with the bed at 250° C, and the plant pressure at 35 atm. The temperature of the gas leaving the reactor, however, began to rise rapidly, and it seemed evident that the dust adhering to the filters was the only active catalyst in the reactor. The run was, therefore, terminated.

After cooling down, the catalyst was easily discharged, leaving a clean and empty reactor. The filters were in a similar condition to that observed at the previous shut-down.

The main operating and performance data are summarized in Table V, and the course of the run is shown graphically in Fig 10. The first twenty days under normal synthesis conditions represent a fairly satisfactory performance, with CO conversions of 95 to 98 per cent, (CO + H_2) conversions of 85 to 90 per cent, and liquid product yields appreciably higher than in Run 1.

The CO conversion plot in Fig 10 illustrates the slow fall in catalyst activity during the first 25 days of the run and the rapid collapse in the last 6 days. The overall decline in catalyst activity throughout the run is reflected in the decrease in the difference between the mean catalyst temperature and the reactor cooling-water temperature. Up to Period 21 this was 34° to 37° C; in Period 31 it was 15° C. The temperature gradient through the catalyst bed, shown by the figures in Table V, appears to be a function of linear gas velocity. Thus, the maximum difference (10° to 12° C) was observed when the linear velocity was low (0·52 to 0·53 ft/sec) and differences of only 7° to 8° C were observed at the higher velocity (0·63 to 0·66 ft/sec). Variations in fluid-bed height are shown in Fig 11. If no withdrawals of catalyst from the reactor had been made during the first twenty-one days of the run, the final bed height would have been of the order of 7·5 ft. Hence, the rate of bed expansion during the later stage of the run was not, in fact, any greater than in the first twenty-one days. The density of the bed was initially 110 lb/cu. ft.; at Period 21, 95 lb/cu. ft.; and at the final shut-down, 80 lb/cu. ft.

The weight and composition of the solids charged to, and withdrawn from, the reactor at various stages throughout the run are given in Table VI. The total carbon formed during the first stage of the run is equivalent to 1·63 g/m³ CO converted, and that found during the last stage, to 1·43 g/m³ converted. These are considerably lower figures than that obtained in Run 1, despite the fact that the mean hydrogen partial pressure in the reactor during the first twenty-one days was lower (167 p.s.i.) than that during Run 1. It would appear that the fresh catalyst used in Run 2 was less active for carbon formation than the "mixed" catalyst used in Run 1. The screen analyses show the anticipated reduction in mean particle size throughout the run (indicated clearly by the increasing proportion of fines, <240 mesh) and confirm previous observations in small reactors [5] that formation of particles larger than the maximum present in the fresh catalyst also occurs. It will be noted that the catalyst found in the overflow receiver was of smaller mean particle size than the main charge in the reactor and that the material collected round the filter banks consisted almost entirely of fines <240 mesh.

The data given in Table VI revealed no obvious cause for the final collapse in the activity of the catalyst. The sulphur contents of the catalyst samples are much higher than would have been expected from the sulphur content of the feed gas as determined intermittently. The final figure of 0·2 per cent sulphur, which is equivalent to an average sulphur content of 0·13 gr/100 cu. ft. of feed gas was not suspected, however, to be sufficient to cause almost complete poisoning of the catalyst. X-ray diffraction analysis of samples of catalyst taken after Periods 8, 22, and 32 and at the end of the run indicated that the main constituent in all cases was iron carbide (Fe_2C, Hägg) and that magnetite (Fe_3O_4) was only present in small amounts. In the later samples some hexagonal Fe_2C and cementite (Fe_3C) was present. The sample taken following the re-reduction treatment before Period 23 contained α-iron, together with some cementite, but the re-reduction carried out after Period 32 and a reduction of the catalyst discharged at the end of the run, carried out at 450° C in the laboratory, produced no change in the X-ray diffraction pattern, *i.e.* the main phase was still Hägg carbide and no α-iron was detectable. There appeared, therefore, to be a link between loss of synthesis activity and stability of the iron carbide to hydrogen at 450° C. It was then found that if a sample of the active

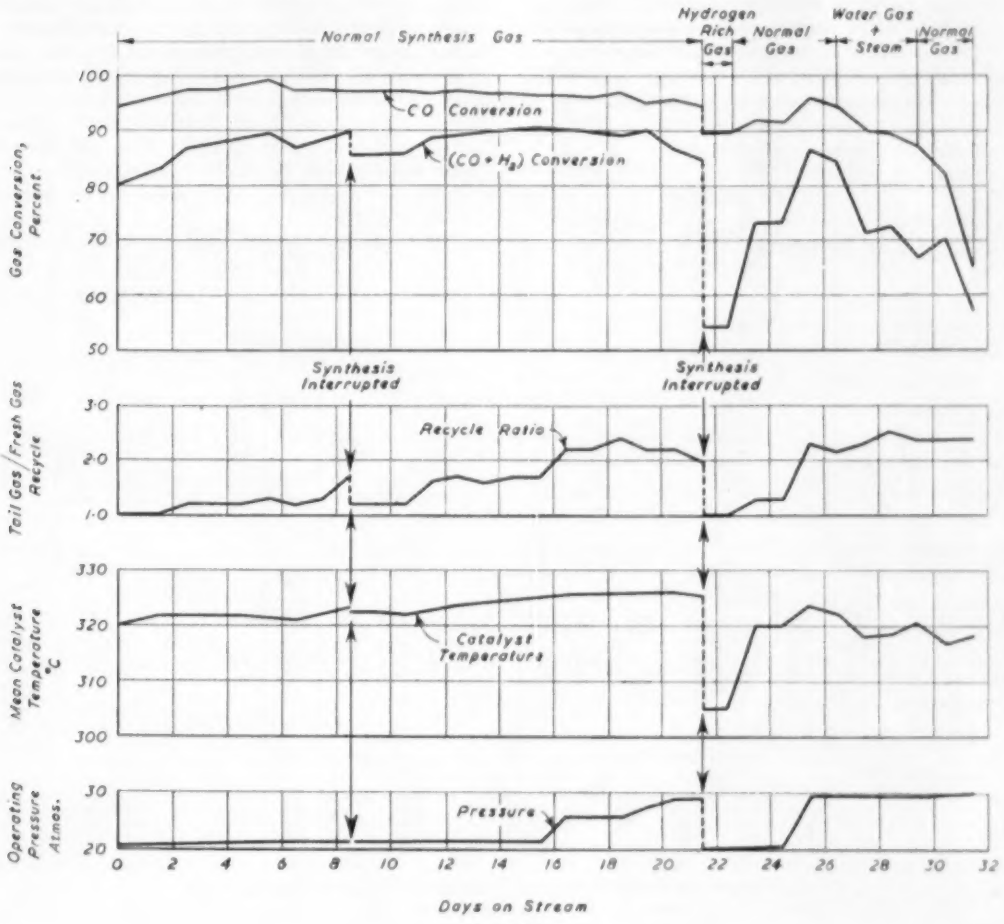


Fig 10

VARIATION IN OPERATING CONDITIONS AND GAS CONVERSIONS WITH TIME—TEST RUN 2

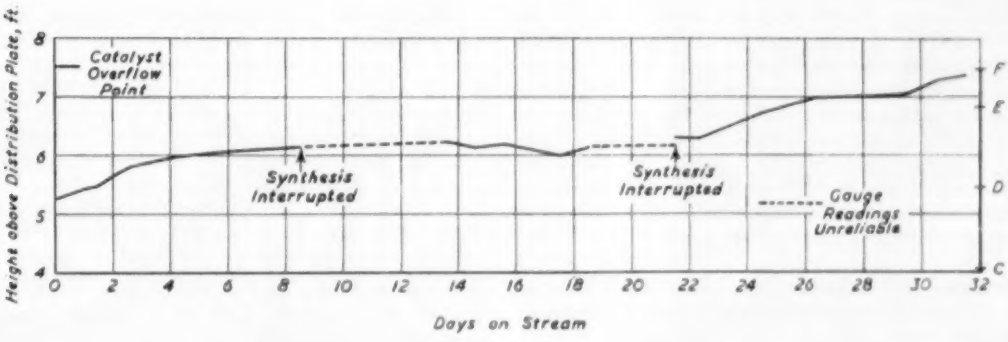


Fig 11

CATALYST BED EXPANSION WITH TIME—TEST RUN 2

catalyst taken after Period 22 was exposed to hydrogen sulphide so that its sulphur content was increased to about 0·2 per cent, the carbide was no longer reduced by hydrogen at 450° C, and its synthesis activity (as during operation under normal synthesis conditions and was similar to that of the product produced in Run 1. Operation with hydrogen-rich gas appeared to increase the average molecular weight of the pro-

TABLE VI

Weight and Composition of Catalyst Charged and Recovered, Test Run 2

Description	Weight, lb	Bulk density, lb/cu. ft.	Benzene-soluble, %	Total iron (in benzene-insol)		Total carbon (in benzene-insol)		Main constituents, % by w (in benzene-insol)			Total sulphur, % by w	Screen analysis, % by w (BS Test Sieves)					
				%	lb	%	lb	Fe_3O_4	Fe_3C	"Free C"		>72	72/100	100/170	170/240	<240	Loss
Catalyst charged to reactor	212	150	—	—	—	—	—	—	—	—	—	1·4	46·2	48·8	2·5	0·7	0·4
Reduced catalyst in reactor at start of synthesis	149	130	—	91·4	136	—	—	—	—	—	0·02	0·8	44·5	50·3	3·4	1·0	Nil
Sample from reactor after Period 6	2·2	121	3·0	75·7	1·6	13·8	0·3	14·2	72·4	6·8	0·03	6·4	36·8	49·2	3·6	4·0	Nil
Catalyst in overflow receiver:																	
Period 12	10·7	130	2·5	73·9	7·6	16·1	1·7	17·1	67·0	9·6	—	3·1	21·0	42·5	7·6	25·5	0·3
Period 16	14·6	130	2·0	73·4	10·5	16·9	2·4	12·0	71·6	10·0	—	3·4	18·4	42·8	4·1	27·3	Nil
Period 21	3·2	128	2·0	74·0	2·3	17·4	0·5	11·6	72·6	10·4	—	6·8	23·0	41·8	6·5	21·5	0·4
Catalyst discharged from reactor after Period 22	155·0	120	2·7	73·0	110·0	17·7	26·7	10·9	72·1	10·7	0·11	8·6	25·3	41·2	7·0	17·5	0·4
Dust adhering to filters	2·0	76	1·0	64·1	1·3	13·1	0·3	62·6	20·7	11·1	—	0·4	0·7	1·5	0·6	95·8	1·0
Total removed	187·7				133·3		31·9										
Catalyst re-charged to reactor	160	121	2·7	73·0	113·5	17·0	26·4	—	—	—	—	—	—	—	—	—	—
Catalyst in reactor after re-reduction in hydrogen	145·6	123	Nil	77·5	112·4	14·1	20·5	6·2	—	—	0·13	—	—	—	—	—	—
Sample from reactor:																	
After Period 32	0·5	103	1·0	69·3	0·4	18·8	0·1	21·5	59·5	13·0	—	—	—	—	—	—	—
After H_2 treatment *	0·3	107	0·5	71·8	0·4	18·6	0·1	12·1	—	—	—	—	—	—	—	—	—
Catalyst discharged from reactor at final shut-down	151·0	108	Nil	71·6	108·5	18·9	28·6	11·7	69·2	12·1	0·20	5·1	24·6	36·6	12·2	21·3	0·2
Dust adhering to filters	2·5	61	0·3	63·1	1·6	29·9	0·7	5·5	65·4	23·6	—	—	—	—	—	—	—
Total removed	154·5				110·9		29·5										
* Add carbon removed by H_2							+ 1·3										
Subtract carbon present initially							−20·5										
Net carbon formed							10·3										

TABLE VII

Properties of Liquid Products (Condensed Oil + Spirit) Test Run 2

Test period	4	8	12	16	18	20	23	27	28	29	30
Synthesis conditions:											
H_2–CO ratio of feed gas	1·99	2·31	2·00	1·96	1·96	1·97	3·78	2·00	1·03 + steam	1·17 + steam	1·16 + steam
Operating pressure, atm, gauge	21	21	21	25	25	29	20	30	30	30	30
Mean catalyst temperature, ° C	322	323	324	326	326	326	305	322	317	317	320
Boiling range, % by w:											
Up to 200° C	82·5	80·7	83·9	84·3	83·8	84·7	72·4	79·2	76·9	74·6	72·3
200°–300° C	12·2	14·1	11·9	10·7	12·9	11·3	13·9	14·2	17·7	17·3	17·0
Residue at 300° C	5·3	5·2	4·2	4·0	3·3	4·0	13·7	6·6	5·4	8·1	10·7
Total Oxygen, % by w (by Unterzaucher method)	5·1	—	6·2	6·8	—	—	8·4	8·5	8·6	—	—
Calorific value, B.T.U/lb (gross)	19,110	—	18,460	18,100	—	18,560	—	—	—	—	—
Analysis of 80°–150° *fraction* *:											
Olefins, % by w	63·8	58·9	53·3	51·6	52·4	49·4	37·2	45·9	47·7	46·1	48·2
OH number, mg KOH/g	54·7	67·2	69·7	72·5	89·2	106·9	173·8	116·1	94·1	88·6	95·4
Acid number, mg KOH/g	20·3	24·4	31·2	34·5	33·3	32·5	17·2	35·1	44·2	50·6	44·1
Ester number, mg KOH/g	15·8	16·7	19·8	20·1	24·3	27·8	24·9	28·4	29·6	29·3	29·4
Total oxygen, %	3·6	4·3	4·9	5·2	5·9	6·5	7·2	6·8	6·9	7·2	7·0

* Approx 30% of total.

determined in a laboratory reactor at 325° C and 30 atm pressure) was extremely low. It was concluded, therefore, that poisoning by sulphur was the main cause of catalyst deterioration in Run 2.

Properties of the liquid products are shown in Table VII. The boiling range showed little variation duct (although this may be due, in part, to the lower catalyst temperature), and the products obtained when using water-gas and steam as feed also contained a higher proportion of high boiling fractions than those obtained with normal synthesis gas. The oxygen content of the product increased with time up to

about Period 23, but at corresponding periods in the run was distinctly lower than in Run 1. There was a roughly parallel increase in the yield of water-soluble oxygen-containing products with time (see Table V). Use of water-gas and steam increased the total quantity of oxygenated products and noticeably increased the ratio of acids to neutral compounds. Use of a hydrogen-rich feed gas, on the other hand, appeared to increase the proportion of alcohols produced.

More complete product distributions for six test periods are given in Table VIII. A carbon and oxygen balance for Test Period 4 is given in Table IX. It will be observed that under normal synthesis conditions appreciably greater specific yields and space–time yields of useful products were obtained than in Run 1; the specific yield reached 160 g/m³ feed gas and the output, 500 lb/day. The proportion of $C_1 + C_2$ hydrocarbons was, for the greater part of the time, less than 20 per cent of the total products, compared with over 25 per cent in Run 1.

In considering the low yields obtained with hydrogen-rich gas in Period 23, it must be remembered that the maximum possible yield of hydrocarbons $[(CH_2)n]$ obtainable with this gas is 127 g/m³, compared with about 202 g/m³ for the normal synthesis gas (H_2–CO = 2, inerts 3·5 per cent). The main effect of using this gas was (as anticipated) an increase in the proportion of light hydrocarbons produced. Thus, $C_1 + C_2$ amounted to over 30 per cent of the total products, and $C_3 + C_4$ to over 50 per cent of the recoverable products.

The results given for Period 29 suggest that there is no advantage to be gained by replacing the normal gas by a mixture of water-gas and steam. The proportion of $C_1 + C_2$ is not reduced, and the distribution of useful products is, on the whole, less favourable than under normal operating conditions.

In a paper dealing primarily with the detailed composition of the products obtained with fluidized-iron catalysts and its bearing on the synthesis reaction mechanism, Weitkamp and Frye [8] give a product distribution relating to operation in an 8-inch dia reactor. This is reproduced in the final column of Table VIII. The precise operating conditions were not disclosed, but were stated to be within the range shown, and were, therefore, comparable with those employed in Run 2. The products shown represented 83·9 per cent of the CO converted, the balance being CO_2. It will be seen that the product distribution is, on the whole, remarkably similar to those for Periods 4, 12, and 17.

To prepare samples of motor fuel for engine test, 2½ kg of average total product from Periods 11, 12, and 13 were washed with aqueous caustic alkali to remove acids, allowed to stand overnight with alcoholic potassium hydroxide to hydrolyse esters, and finally washed repeatedly with water. This

TABLE VIII

Product Composition and Yields for Selected Periods of Test Run 2

Test period	4		12		17		21		23		29		Data of Weitkamp and Frye [8]
Synthesis conditions:													
H_2–CO of feed gas	1·99		2·00		1·83		2·01		5·78		1·17 (+ steam)		1·8
Space velocity, vol/vol fluid catalyst/hr	1330		1330		1330		1330		1170		900		7–9 cu. ft. CO/lb Fe/hr
Mean catalyst temperature, °C	322		323		326		325		305		317		280–360
Operating pressure, atm, gauge	21·1		21·5		25·6		29·0		20·3		29·5		18–38
	g/m³ feed gas	% of total	g/m³ feed gas	% of total	g/m³ feed gas	% of total	g/m³ feed gas	% of total	g/m³ feed gas	% of total	g/m³ feed gas	% of total	% of total
Total products:													
Methane	20·8	10·8	18·9	9·8	19·6	9·7	24·1	12·6	20·6	17·6	19·9	11·3	10·7
Ethylene	10·0	5·2	9·5	4·9	10·0	5·0	9·0	4·7	8·8	7·5	14·3	8·1	4·7
Ethane	8·6	4·5	7·7	4·0	10·2	5·1	9·2	4·8	6·6	5·7	8·1	4·6	5·7
Propylene	23·2	12·1	21·4	11·1	28·5	14·1	19·5	10·2	14·5	12·4	26·3	15·0	11·8
Propane	3·4	1·7	3·4	1·7	1·6	0·8	2·8	1·5	3·8	3·4	4·0	2·3	3·2
Butylenes	15·8	8·2	18·0	9·3	20·8	10·3	12·3	6·5	21·0	18·0	18·8	10·7	9·3
Butanes	2·7	1·4	3·0	1·5	0·5	0·3	2·3	1·2	3·6	3·1	3·8	2·1	3·1
C_5 and higher	86·8	45·2	83·0	42·9	82·8	41·1	78·1	40·9	23·1	19·8	50·7	28·8	39·4
Acids in reaction water	6·0	3·1	7·0	3·6	7·6	3·8	8·1	4·2	3·5	3·0	12·9	7·4	5·4
Neutral organic matter in reaction water	14·7	7·8	21·4	11·2	19·7	9·8	25·7	13·4	12·3	10·5	17·1	9·7	8·7
Total	192·0	100·0	193·3	100·0	201·3	100·0	191·1	100·0	116·8	100·0	175·9	100·0	100·0
Recoverable "useful" products:													
Propane and butanes	6·1	4·0	6·4	4·1	2·1	1·3	5·1	3·4	6·4	7·9	7·8	5·8	8·0
Propylene and butylenes	39·0	25·6	39·4	25·1	49·3	30·5	31·8	21·4	35·5	44·0	45·1	33·8	26·8
Crude naphtha, E.P. 200° C	71·5	46·8	69·8	44·4	69·6	43·1	66·2	44·5	16·7	20·6	37·9	28·4	50·0
Middle oil, 200°–300° C	10·6	6·9	9·9	6·3	9·8	6·1	8·8	5·9	3·2	4·0	8·8	6·6	
Heavy oil and wax	4·7	3·1	3·3	2·1	3·4	2·1	3·1	2·1	3·2	4·0	4·0	3·0	
Acids and neutral chemicals	20·7	13·6	28·4	18·0	27·3	16·9	33·8	22·7	15·8	19·5	30·0	22·4	15·2
Total	152·6	100·0	157·2	100·0	161·5	100·0	148·8	100·0	80·8	100·0	133·6	100·0	100·0
Output of "useful" products:													
lb/day	467		492		502		465		227		324		—
lb/cu. ft. fluid catalyst/day	304		312		317		295		141		180		—
gallons/day	66		70		71		66		32		46		44 (oil)

treatment effected a loss in weight of 17·5 per cent. The dried material was then distilled and the fraction distilling below 200° C divided into two equal portions. One portion was subjected to vapour-phase treatment at 350° C over activated alumina (Peter Spence and Sons Ltd., Type " A ") at a liquid space velocity of 0·6 vol/vol/hr. As a result of this treatment the hydroxyl number fell from 31·5 to 1·5 mg KOH/g and infra-red analysis showed an increase in internal double bonds at the expense of terminal double bonds. The octane number (CFR Research Method) of the untreated naphtha was 73·8 and of the alumina-treated material, 74·9.

TABLE IX

Carbon and Oxygen Balance for Period 4, Test Run 2

(Basis: 1000 litres feed gas)

Input			Output		
	Carbon, g	Oxygen, g		Carbon, g	Oxygen, g
Carbon monoxide	175·8	234·2	Carbon monoxide .	3·3	4·4
Carbon dioxide .	8·0	21·4	Carbon dioxide .	25·2	67·2
			Hydrocarbons C_1–C_4	69·8	—
			Condensed oil and spirit . . .	70·6	4·5
			Water soluble acids and other compounds . .	12·2	7·2
			Water . . .	—	170·5
			Total . .	181·1	253·8
			Loss . .	2·7	1·8
Total .	183·8	255·6	Total . .	183·8	255·6

THERMAL DATA

The temperatures at various positions during Period 27 of Test Run 2, which are typical of those obtaining in the high-pressure reactor system, are given in Fig 12.

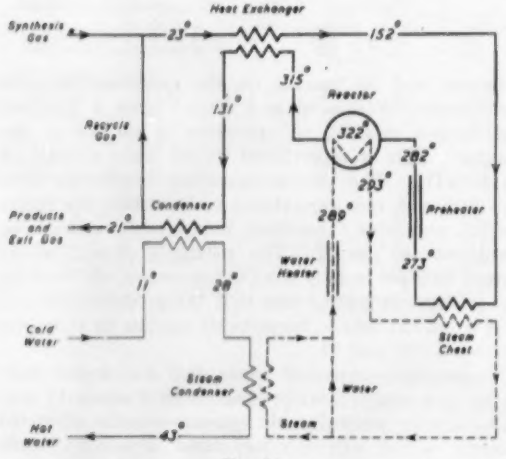


FIG 12

PILOT SCALE FLUID CATALYST HYDROCARBON SYNTHESIS PLANT TYPICAL TEMPERATURES IN REACTOR SYSTEM, °C.

Calculations of the heat entering, circulating within, and leaving the system were made for this period, and the results are expressed diagrammatically in

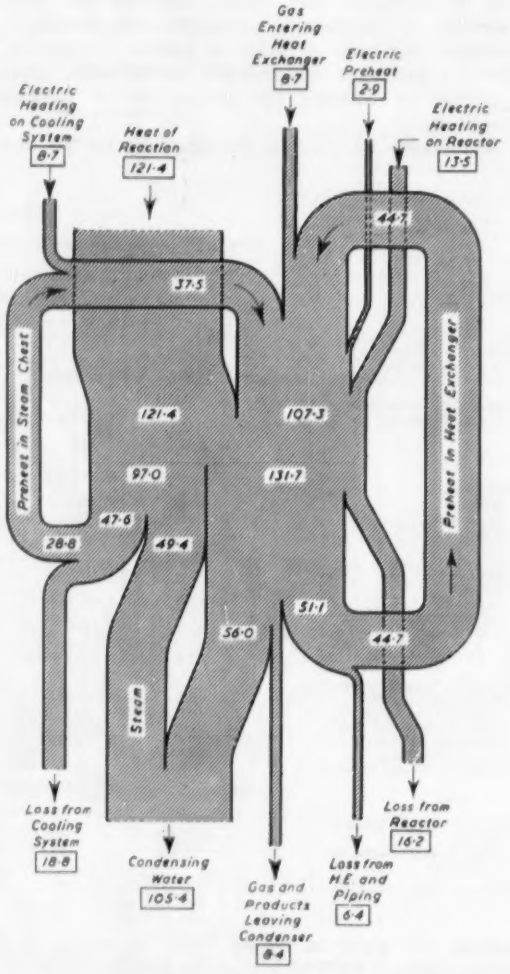


FIG 13

HEAT FLOW DIAGRAM. SENSIBLE AND LATENT HEAT. THOUSANDS B.T.U/HR. DATUM 32° F

Fig 13. The figure of 121,400 B.T.U/hr for the heat of reaction was obtained by difference.

The heat of reaction was calculated for Period 4, Test Run 2, from mass flows and calorific values, as follows:

Input:	(CO + H_2) 59·81 lb/hr . .	681,400 B.T.U/hr
Output:	Gas (CO,H_2,C_1,C_2) 7·13 lb/hr (27·7 per cent of input) .	194,900 B.T.U/hr
	Recoverable products (C_3,C_4, oil, spirit, and water-soluble products) 19·52 lb/hr (53·7 per cent of input) . .	365,500 B.T.U/hr
	Total	560,400 B.T.U/hr
Heat of reaction, by difference . .		121,000 B.T.U/hr

For the conversion of 1771 cu. ft/hr ($CO + H_2$), this represents a heat evolution of 68·4 B.T.U/cu. ft. ($CO + H_2$) converted. On the basis of this figure, the heat of reaction in Period 27, for a conversion of 1755 cu. ft. ($CO + H_2$)/hr, was 120,000 B.T.U/hr.

To obtain a figure for the coefficient of heat transfer from the catalyst to the cooling water, a separate thermal balance for the cooling system was worked out as follows :

Input :	Electric heating on cooling system		8,700 B.T.U/hr
Output :	To gas in steam chest .	37,500	105,700 B.T.U/hr
	To water (steam condenser) . . .	49,400	
	Loss to atmosphere .	18,800	
Giving heat from catalyst, by difference .			97,000 B.T.U/hr

As the external area of the tube bundle is 19·0 sq. ft. and the temperature difference, catalyst to water, 52° F, the transfer is 98 B.T.U/sq. ft/° F/hr. This figure is of the order expected for a bed of fluidized iron catalyst at the linear gas velocity, 0·54 ft/sec, used.

COMMENTS ON PLANT PERFORMANCE

It is believed that the reactor was designed for too low a linear velocity and that this tended to exaggerate the relatively steep temperature gradient through the catalyst bed. The tubular filters proved entirely satisfactory, and the pressure-drop across them remained negligible without blowing back. However, regular blowing back on a rigid, short time cycle appears to be essential to prevent catalyst accumulating on the filters and causing local overheating.

When designing the plant, it was anticipated that wax would not separate from the oil in the condenser. The reduction of cooling efficiency which was experienced, due to a wax film building up on the tube surfaces, indicates that a wax by-pass, as fitted during Run 2, is necessary.

It is unfortunate that trouble was experienced with the recycle pumps. This type of pump was chosen on the strength of the good service obtained from several similar but smaller pumps used on the laboratory-scale plants. The larger pumps were the first to be made for conditions comparable with those in the pilot plant. The main fault was traced to excessive pressure inside the diaphragm chamber, which caused low efficiency and overloading of the Hycar diaphragms and the variable stroke mechanism. Modifications made during the runs and subsequently, by the makers and Fuel Research staff, will, it is believed, much improve their efficiency and life.

The steam leak which persisted for a considerable part of Test Run 2, and necessitated frequent pumping of water into the reactor cooling system, resulted from the blowing out of asbestos jointing material which had been inserted in error. The joint is designed to be metal-to-metal contact between a " nibbed " and a flat face. The leak was reduced at intervals by tightening the bolts, but finally became excessive, as the escaping steam scored the metal, and a shut-down was necessary to allow the surfaces to be refaced.

Ermeto pipe couplings, including those holding up to 85 atm steam pressure, proved satisfactory. The valves on the hot gas and product lines were difficult to keep leak tight when opened and shut frequently.

Apart from the first 72 hours of Run 1 and the first 24 hours of Run 2, when a chemist was present on each shift, the plant was operated without difficulty by skilled labourers (2 per shift) experienced in the operation of the small-scale synthesis units.

COMPARISON OF THE RESULTS WITH THOSE OBTAINED IN A SMALL REACTOR

The gas conversions obtained in both pilot plant runs were lower than had been anticipated from experience in the 1-inch dia reactor, previously described,[3] and it was only in Run 2 that the product distribution was of the expected type. It was of interest, therefore, to carry out experiments in the 1-inch dia unit with samples of catalyst withdrawn from the pilot-plant reactor. After Run 1, experiments were carried out with a sample of reduced catalyst taken from the reactor before starting synthesis and with an average sample of the catalyst discharged at the end of the run (after Test Period 15). The results are shown in Table X, together with those of a run with the fresh catalyst which formed part of the charge for Test Run 1. (A run in the empty

TABLE X

Comparison of Results Obtained in Test Run 1 with those Obtained in 1-inch Diameter Reactor

Plant	Pilot plant (Run 1)		1-inch diameter reactor		
Catalyst	Catalyst batch PB3 + used catalyst from trial run		Reduced catalyst from pilot plant	Spent catalyst from pilot plant	Batch PB3 (reduced in 1-inch unit)
Test period	10	15	10	1	10
Mean catalyst temperature, ° C	325	322	325	322	325
Operating pressure, atm, gauge	20	20	20	20	20
H_2–CO of fresh feed	1·87	1·92	2·00	1·98	1·96
Throughput, cu. ft/lb Fe/hr	18	19·7	16·2	26·7	19·6
Recycle ratio, tail gas/fresh feed	2·4	2·5	3·0	2·0	3·0
CO converted, %	97·1	90·8	98·7	96·8	99·4
($CO + H_2$) converted, %	92·8	85·7	95·6	92·8	97·8
CO converted to CO_2, % of total	9·1	7·8	1·4	6·0	3·7
CO converted to methane, % of total	10·0	11·1	13·2	12·7	8·2
CO converted to higher hydrocarbons, % of total	80·9	81·1	86·4	81·3	88·1
Yields, g/m³ feed gas :					
Condensed oil and spirit	66·1	42·5	74·9	70·6	92·3
Organic matter in water	26·2	*ca.* 30	31·6	25·5	21·8
Boiling range of condensed oil and spirit, % by w :					
Up to 200° C	86·9	81·6	83·6	88·3	71·7
200°–300° C	9·6	11·5	12·5	8·7	14·8
Residue at 300° C	3·5	6·9	3·9	3·0	13·5
Properties of 80°–150° C fraction :					
Olefin content, %	51·7	35·0	57·8	50·7	60·7
OH number, mg KOH/g	127·7	216·8	94·0	165·6	55·7
Acid number, mg KOH/g	36·4	42·8	34·0	53·0	33·4
Ester number, mg KOH/g	68·0	79·4	53·0	33·0	17·5
Total oxygen, % by w	9·6	13·2	7·7	9·7	4·5

1-inch reactor at a pressure of 30 atm and a temperature of 330° C showed no volume contraction or change in the gas composition.) It was not always possible to reduce the recycle ratio in the small unit to as low a value as that obtaining in the pilot plant. Hence comparison of catalyst activity in the two units must be based mainly on the CO conversion, which is not sensitive to changes in recycle ratio. The results show that a particular sample of catalyst exhibits a definitely higher activity in the small unit than in the pilot plant, the difference being particularly marked with the catalyst discharged at the end of the pilot plant run. The results also suggest that the product distribution obtained in Test Run 1 was a function of the mixed catalyst employed and not one of plant scale or geometry. The product distribution obtained

TABLE XI

Comparison of Catalyst Activities in Pilot Plant (Test Run 2) and 1-inch Diameter Unit

Plant	Pilot plant	1-inch unit		Pilot plant	1-inch unit	Pilot plant	1-inch unit	1-inch unit	
Catalyst		From pilot plant after initial reduction			From pilot plant after Period 8		From pilot plant after Period 22	From pilot plant after final shut down	As previous column, after re-reduction in 1-inch unit
Test period	2	1	4	8	1	21	1	1	1
Mean catalyst temperature, ° C	322	323	323	323	323	325	325	325	325
Operating pressure, atm, gauge	20	20	20	20	20	29	28	30	30
H_2–CO of fresh feed	2·08	1·82	1·82	2·31	2·33	2·10	1·91	1·83	1·80
Feed gas rate, cu. ft/lb Fe/hr	15	16·0	15·6	16	18·7	19	23·3	24·6	24·7
Recycle ratio, tail gas/fresh feed	1·2	3·8	1·5	1·7	5·8	3·2	5·5	5·5	5·5
CO converted, %	97·5	99·5	99·5	97·3	97·9	94·8	96·1	23·1	22·9
(CO + H_2) converted, %	86·9	97·7	94·9	90·0	91·9	84·9	91·7	17·8	17·9
Usage ratio, H_2–CO	1·75	—	1·78	2·07	2·12	1·70	1·72	—	—
CO converted to CO_2, % of total	9·2	—	4·3	8·4	7·3	9·8	5·1	—	—
CO converted to CH_4, % of total	10·5	—	8·3	7·8	13·5	9·3	7·8	—	—
CO converted to higher hydrocarbons, % of total	80·3	—	87·5	83·8	79·2	80·9	87·1	—	—

with the fresh mill-scale catalyst in the 1-inch unit was broadly similar to that obtained with fresh catalyst in the 2nd Test Run in the pilot plant.

In two runs in the 1-inch dia unit with freshly reduced mill-scale catalyst operated under conditions similar to those used in Test Run 1, no change in CO or (CO + H_2) conversion was obtained in a period of 15 days.

Data showing the activity in the small unit of samples of catalyst withdrawn from the pilot plant at various stages of Run 2 are reported in Table XI. The freshly reduced catalyst and that taken from the pilot plant after Period 22 show a significantly higher activity in the 1-inch unit. The activity of catalyst taken after Period 8 is, however, little different from that observed in the pilot plant during Period 8. The extremely low activity in the small unit of the catalyst recovered after the final shut-down is qualitatively in agreement with experience in the pilot plant, as is also the complete failure of re-reduction in hydrogen to effect any improvement in activity.

In the light of all the data in Tables X and XI it must be concluded that under a particular set of operating conditions a batch of catalyst will give a significantly better performance in the 1-inch dia unit than in the 8-inch dia pilot plant. In view of the fact that the height–diameter ratio of the fluid bed is of the order 25 to 30 in the 1-inch unit and 10 in the pilot plant, this would have been expected from the work of Askins, Hinds, and Kunreuther [9] and that of Gilliland and his co-workers [10, 11] on gas mixing in beds of fluidized solids. These workers concluded that owing to the approach to conditions of complete gas mixing in fluidized beds the extent of conversion of gaseous reactants will be less than in a fixed bed and will decrease with decrease in the height–diameter ratio of the fluid bed. Danckwerts,[11] however, concluded from experiments in a fluid cat-cracker regenerator that conditions approached more nearly to those of piston-flow than to those of complete mixing. Whether or not gas mixing is the explanation, there is experimental evidence that reaction rates in fluidized beds are sensitive to height–diameter ratio.

Gilliland, Mason, and Oliver [12] studied the oxidation of NO to N_2O_4 in a fluidized-bed system and found that, with H–D = 10·4, the conversion obtained was 95 per cent of that obtained with H–D = 27, and with H–D = 6 it was only 85 to 89 per cent. Hoog [13] found that in fluid catalytic cracking the percentage conversion of the feed decreased with increase in bed diameter for a given catalyst, temperature, and throughput, the effect becoming more marked with increase in the level of conversion.

CONCLUSIONS

Although the runs were marred by mechanical troubles, they established the soundness of the main design features of the plant and achieved the desired objectives. Periods of up to 350 hours continuous operation with high gas conversions and product yields

and with a low rate of carbon formation were achieved. The product distribution was apparently fairly typical of that obtained in American "Hydrocol" pilot plants. The results suggest that with a feed gas containing an average of 0·13 gr total sulphur per 100 cu. ft. the life of the mill-scale catalyst, with the conditions used, is of the order of 30 days (or about 1 lb catalyst per 100 gal recoverable product). Further experiments using more highly purified gas and operating with continuous catalyst replacement would be required to establish reliable data for catalyst consumption.

The observed change in the oxygen content of the products with time is clearly a function of catalyst ageing in a "batch" operation, and if operated with continuous catalyst replacement (or regeneration) the catalyst in the reactor would reach a steady-state condition, and products of constant quality should be produced.

The experiments provided the type of thermal and other data required for the design of a full-scale plant, but the comparison of the synthesis performance obtained in the pilot plant with that obtained in the small unit, taken in conjunction with published work on the characteristics of fluid–solids systems, suggests that it would be unwise to design a full-scale reactor by a simple increase in dimensions of the pilot-plant reactor. It appears that if a full-scale Fischer–Tropsch reactor using a dense-phase fluid catalyst is to reproduce the performance obtained in a pilot-scale reactor, the height–diameter ratio of the bed must remain unchanged. This could be achieved by using, on the full scale, a multiplicity of relatively narrow tubes (*e.g.* 2 to 4 inches dia), containing the fluid catalyst, within a single pressure-resisting shell. Such an arrangement has been patented by Keith [14] but without reference to the point discussed above.

The experience gained in the operation of this pilot plant will be of value in planning and carrying out similar work with a reactor designed for the slurry process. This process, which is considered to be of greater promise for the Fischer–Tropsch synthesis than the dense-phase fluidized process, will be the subject of more extensive pilot-plant work than that described in this paper.

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DISCUSSION

D. P. Plummer: There is no need for me to emphasize the importance of the study of fluidized catalysts; indeed, it is not an exaggeration to say that the technique of fluidizing catalysts has changed the whole face of the petroleum industry. In this paper, not only is the importance of fluidizing catalysts emphasized but the actual fundamental interest in the phenomena is also apparent.

This stirs the philosopher within one to contemplate the "middle way" in which one has a velocity higher than that sufficing for percolation and lower than that which is necessary for entrainment. Thus, one studies this defined range in which these fascinating phenomena occur. First, however, to anyone whose primary interest and education has been in petroleum, there is a very great evidence of contrast between this work and the application of fluidized catalysts to catalytic cracking, in which the main interest is in operation at low pressure. Here, it is in operation at very high pressure. In catalytic cracking one considers endothermic reaction; here the reaction is quite strongly exothermic. In catalytic cracking there is an expansion—the products occupy more room than the reactants; here there is actually a contraction during the reaction. In catalytic cracking, one considers the properties of a light catalyst, of a fluidized density much less than that of water; here it is distinctly heavier than water. Also in catalytic cracking, one considers rapid reaction and very short catalyst life; this is comparatively slow reaction and a very long catalyst life. Also, in catalytic cracking, one is, as the speaker emphasized, content with a moderate conversion; 60 per cent is considered quite high, and 70 per cent very

high indeed. The speaker this evening, as you may remember, positively disdained conversions of 90 per cent, and was not content until he had 96 to 99 per cent.

This persistent and continuous contrast emphasizes the universality and the persistence of the phenomenon of fluidization, and that it is indeed a fundamental property of matter.

In regard to the paper, a great number of questions had occurred to me, but only very few remain after Dr Hall's presentation.

What is the actual shape of these particles? Before reduction, they were, I imagine, just loose irregular flakes, but afterwards, and particularly after 14 or 18 days use in a reactor, did the shape change; did they become more spherical; or what was the shape of them?

I cannot quite persuade myself that the mass was truly fluidized. The linear velocity is what is commonly used for pilot plant studies, but it seemed a little on the low side at ½ ft/second. If the bed is truly fluidized, then what is going to the bed is not so much the reactants as the products, and if there is a contraction of 75 per cent, then the actual linear velocity in the bed is only a quarter of that—which seems rather low. If one considers the mass velocity, the weight of the stuff going through per weight of catalyst per hour, a figure of 0·3 or 0·4 could be calculated, and that seems a little on the low side.

Then also the characteristic of a fluidized bed is a uniform temperature. Well, we have heard of an actual drop of temperatures of the bed. Is that a consequence of an exothermic reaction, being balanced perhaps by over-sized heat transfer equipment? There is a high heat transfer coefficient and that is a characteristic of fluidization, but fluidization near to the actual bundle of pipes in the centre of the bed would have been sufficient to have obtained that. Was all the bed uniformly fluidized? One of the characteristics very simple to measure, and I am sure the authors must have observed it, is the actual variation of pressure drop through the bed with the velocity. Did they ever observe this constancy of pressure drop over a range of velocities, that is characteristic of a fluidized bed?

The author has some misgivings on this himself, for he says he wishes he had designed the reactor for rather a higher velocity. Surely it could have been decided. If he had made the 30 atm run, say at 19, he could have had nearly twice the velocity, possibly 0·8, and that would have been very interesting. I would like the author's opinion on that point of view, particularly for this dependence on L over D ratio. Would it have been true to say that if the velocity had been higher, the fluidization would have been more complete, and then possibly dependence on the L over D ratio would not have been observed.

I cannot persuade myself that a fluidized bed of catalyst has diameter. A heap of sand has diameter, but the fluidizing process is concerned with the properties of a liquid. Now, there is no such thing as the diameter of water, for if you put it in a container of a certain diameter, the diameter of the container cannot be said to entirely perpetuate itself in the properties of the liquid inside it. I am not at all certain therefore that a completely fluidized bed has diameter; whether it has length is perhaps more arguable, but it should not have an L over D ratio—it should not have a dependence on such dimensions of the container, there should be sufficient back mixing to obviate this. I wonder, had the velocity been higher and the fluidization more perfect, if this trouble of L over D ratio would have occurred, and would it have been necessary to say that the full-scale reactor must be a mass of 4-inch tubes?

This is not necessarily a criticism of the paper, because Dr Hall has been extraordinarily honest in his presentation. He is not afraid to give the most ordinary explanations of apparent discrepancies. But what exactly is the point of all this work? There are two ways of looking at it. You can say, "I will apply the classical state of fluidized bed to this reaction and see what occurs." Or you can say to yourself, "I know perfectly well what this reaction wants, it has a long catalyst life, it has the dependence on L over D ratio, but it gives out so much heat that I cannot use a fixed bed, I cannot get rid of the heat, therefore I will fluidize the bed just enough to get rid of the heat but I will perpetuate the other characteristics of the fixed bed." Has the author really been as subtle as this? It is a question that I would only ask of a very honest man. Is the answer to this research that bad fluidization is really what the authors wanted to achieve?

Dr C. C. Hall: The shapes of the original catalyst particles, as seen under the microscope, are irregular, but after quite a short period of synthesis, the particles are almost spherical, and fluidize more readily. The unreduced mill-scale, with a bulk density of about 2·4, has a critical fluidizing velocity of about 0·3 ft/second, and it fluidizes quite vigorously at ½ ft/second. After reduction, its bulk density is down to about 1·9 or 2·0, and after some days of synthesis is reduced still further; it should therefore have been quite adequately fluidized in the pilot plant. Certainly, the bed obeyed all the usual criteria; the pressure difference across the bed was not sensitive to change in velocity, and if the bed had been only partly fluidized, the plots of pressure difference against depth of bed would not have been linear. We are quite convinced that our catalyst bed was completely and uniformly fluidized in Test Runs 1 and 2.

Then there is the question of the effect of gas contraction on the linear velocity. The actual contraction across the reactor was probably about 10 per cent, because we were re-circulating one-and-a-half to two volumes of tail gas, and all the products were in the vapour phase, so that although the overall contraction (feed gas–cold tail gas) was about 75 per cent, the actual volume contraction across the reactor was negligible from the point of view of the effect on linear velocity.

Mr Plummer is perturbed by the observed temperature gradient in the bed. This is to some extent a function of height–diameter ratio; as the ratio increases, one approaches more nearly to conditions in a fixed bed, and if one carries out this exothermic reaction in a fixed bed at high throughputs, one gets a very steep temperature gradient falling in the direction of gas flow. P. C. Keith has stated that the CO concentration at the inlet of a fluidized reactor must not be more than about 10 or 12 per cent, otherwise one gets an excessively high temperature at the gas inlet. In a bed 20 ft dia by 10 ft deep, one would hardly expect to find a vertical temperature gradient, but I think one would certainly expect one in a bed 6 ft deep by 8 inches dia.

We obtained some evidence that the temperature gradient was reduced by an increase in linear velocity, but we did not wish to alter our conditions (operating pressure or recycle ratio), permanently, to achieve a higher linear velocity.

On the question of whether the fluid bed has a diameter or not, I think you have got to bring the retaining walls into consideration, and as you move those containing walls farther apart the lateral gas diffusion increases. This was the basis for Askins' explanation of the increase in the degree of approach to complete gas mixing, as the bed diameter is increased.

Finally, I appear to have been asked the "sixty-four dollar question"—what was the object of this work?

It was, of course, to determine whether the dense-phase fluid-catalyst synthesis process was operable on a pilot plant scale, and to see if there were signs of plant size and geometry factors which would affect the translation of the process to the full scale. We feel, even on the basis of the admittedly incomplete work we did, that the dense-

phase fluid-bed Fischer–Tropsch synthesis is definitely operable, but that it would be unwise to scale-up from the pilot plant by a simple increase in reactor size. I rather feel, however, that there was more to this question than I have answered; perhaps Mr Plummer could amplify it?

D. P. Plummer: What I meant to say was this: I did have some suspicion as to whether the bed was completely fluidized, not in the vertical sense, but in the lateral sense. You could have complete fluidization up the middle but an almost quiescent state in an annular portion of the bed. Supposing you did achieve that state, is the explanation simply that you wished to do so? What I meant to question was the attitude to the phenomena, do you say to yourself, "I wish perfectly to fluidize the bed, and see how it will affect the reaction," or do you say, "I know what the reaction requires and I think it needs imperfect fluidization, just enough to get rid of the heat of reaction."

Dr C. C. Hall: Certainly we had every intention of having the bed completely and uniformly fluidized, and I do not think there is any doubt that we achieved it.

It is most unlikely that we had a quiescent state in the outer annulus, because as a result of having the 2-inch diameter cooling tube in the centre of the bed, the main reaction space was, in effect, an annulus.

M. H. M. Arnold: The existence of a temperature gradient in a fluidized bed is fully consistent with perfect fluidization, for fluidization does not imply vertical mixing of the burden. In a pilot plant for phthalic anhydride, using a fluidized bed of supported vanadium pentoxide catalyst, with a bed area of about 1 m^2, annular in form with an annular width of 15 cm, cooled countercurrently with air jackets, and a bed depth of 3 m, differences of 40° C between top and bottom can be obtained with a gas velocity of 10 cm/sec (critical velocity 8 cm/sec). On increasing the linear velocity the differential falls, but could never be reduced below 5° C at 25 cm/sec. There can be no question of imperfect fluidization for, with such a highly exothermic reaction (about 5000 Tcal/ton product), any failure to fluidize leads at once to hot-spotting and coking. An even better example of the fact that mixing is not required is the fluidized recovery of ethylene from coke-oven gas by sorption on charcoal (B.P. 569,691, worked very efficiently on the semi-technical scale), where mixing must be completely prevented in order to obtain a sharp adsorption front.

It is in fact a pity that so many workers regard fluidization as a technique that necessarily implies mixing.

Dr C. C. Hall: That is very interesting and fits in well with our observations. The naphthalene oxidation is about 30 per cent more exothermic than the Fischer–Tropsch reaction, so that one would expect the rather more magnified temperature-difference effect that Mr Arnold observed.

I also agree that despite the temperature gradient the system is quite stable. We had no hot spots, and at the conclusion of the experiments the catalyst ran out smoothly through a ¼-inch dia hole at the bottom of the reactor, with no deposits, coke, or anything of that sort.

P. J. Grey: I would like to ask three questions which are inter-related:

First, Dr Hall, do you believe that the breakdown of your catalyst particles was due to attrition, or due to the exothermic reaction which was being carried out on the surface of the catalyst? Together with that, did you investigate at all the effect of particle size distribution on the stability of the fluidized bed, that is, was the bed free from channelling and slugging with various particle size distributions; and, finally, you have used sintered brass filters to remove the fines from your exit gas. In an industrial plant, presumably, you would have to rely principally on cyclones or some other means of that type. Did you carry out any studies on the elutriation of fines from your fluidized bed?

Dr C. C. Hall: The cause of breakdown of particles is almost entirely due to chemical reaction in the sense that carbon is formed by the reaction $2\,CO \rightarrow CO_2 + C$ within the core of the particle and causes expansion and disintegration. It will be aggravated by local overheating, since this reaction is favoured by high temperatures. We fluidized a bed of mill-scale catalyst for about 10 days at reaction temperature in hydrogen and found a negligible amount of breakdown by attrition.

We had no evidence of channelling or slugging at any time in the pilot plant reactor, and tests in glass models indicated that the spent catalyst, with its greatly extended particle-size range, fluidized rather more smoothly than the fresh catalyst.

We should, of course, use cyclones in a large plant, but it is more convenient, and I think usual practice, on plants of this size to use filters.

J. P. Asquith: I should like to ask the authors whether, from their experience of fluidized-bed technique, they have any improvements from the chemical engineering viewpoint which they would wish to incorporate.

A. H. Taylor: There are two ways in which we might improve the reactor. We would slightly decrease the diameter in order to increase the linear gas velocity, as already mentioned. We put rather more cooling surface than we would do again. When designing the plant, we assumed a heat transfer figure of 72 B.T.U/sq. ft/° F/hr, whereas we got 98. A smaller cooling surface would allow a greater temperature difference between the catalyst and the cooling system, and hence a lower steam pressure. As regards other parts of the plant, if we were starting again, I think we would use larger pipelines in one or two places, and increase the size of the heat exchanger somewhat, otherwise I do not think we would make any appreciable alteration.

There is, however, the question of blowing back the filters, which as stated in the paper, must be done regularly.

R. J. Waugh: Those concerned with the design and operation of fluid bed units are familiar with the problems associated with erosion caused by catalyst movement. Although the unit has run for a relatively short period, could the authors say whether any evidence of erosion was found within the reactor, with particular regard to the cooling coils.

Dr C. C. Hall: I believe, in fluidized systems, the most serious erosion takes place in slide valves and transfer lines. In our case there was no circulation of catalyst, and, except for a slight polishing, the effect on the cooling tubes was negligible. The total number of days on stream was only about 50 to 60, so that I do not think one would expect to see very serious erosion on the internal surfaces of the reactor.

Dr A. S. Freeborn: I am not certain whether my interpretation of the figures for the operating conditions was correct. I did not follow how it was possible to obtain so high a conversion when the recycle ratio, tail gas over fresh gas, was greater than one.

Dr C. C. Hall: The CO conversion is based on the fresh feed gas and is virtually independent of recycle ratio. The hydrogen conversion (and hence the total CO + hydrogen conversion) increases with increase in recycle ratio due to suppression of the water-gas shift reaction.

C. A. Plummer: How is the small reactor temperature controlled, that is, how is the heat of reaction removed; and what was the temperature differential over the small reactor?

Dr C. C. Hall: The small reactor was a simple one-inch dia tube, about 6 ft long with a winding direct on the wall, and we relied for cooling on heat loss to atmosphere. We did get temperature gradients of about 10°, that is to say, the temperature fell by about this amount from gas inlet to gas outlet.

C. A. Plummer: Does one obtain a greater temperature differential in a small reactor operated at perhaps the same average temperature as a pilot plant, and if so, would this explain the higher conversion. This might be expected in view of the higher L over D for the small reactor, which should cause it to behave more like a fixed bed reactor.

Dr C. C. Hall: The temperature gradient may have been somewhat greater in the small unit, about 10° rather than the 7° to 8° in the pilot plant.

Lt.-Col. S. J. M. Auld: I would have liked to ask the sixty-four dollar question myself, because we have undoubtedly had a complete analysis of a condition of design and operation, despite the fact that it is one of synthesis. Indeed, the paper has been much more a study of fluidization than that of the Fischer–Tropsch synthesis. We are inclined these days, thanks largely to the Middle East oil production, to overlook the importance of synthesis. Even in the U.S.A., where large scale work was embarked on to see whether this would be a future method of ensuring supplies of high cetane value diesel fuel, it has gone into the background.

In that connexion I have been wondering to what extent Germany was studying means of improving the operation of synthesis by similar methods to that we have been listening to tonight. For that reason perhaps Dr Hall will tell us whether the Germans have been working on similar lines, but more from the industrial point of view rather than the basic study of fluidization, which has been the mainspring of the Greenwich work. If there is any work of that kind going on, we would like to hear of it.

Dr C. C. Hall: It is really German work which has formed the basis of the process which, from our small-scale experiments, we think is the most promising, that is, the liquid-phase suspension process. The Germans were quite aware that the original fixed-bed process was probably not the most suitable way of carrying out the synthesis, and throughout the war years they, were working on other systems. Although they did not make any study of fluidization as applied to Fischer–Tropsch, they were very active in the field of slurry synthesis, the system where the finely-divided catalyst is suspended in a liquid medium and where, again, you get very good heat transfer from the catalyst to cooling coils immersed in the slurry, and with the advantage over the fluidized system of much greater selectivity, that is, a higher proportion of one's CO and hydrogen is converted into the desired products.

They have also made great improvements in their fixed-bed process by using gas re-circulation; and an improved fixed-bed process with a very much higher output per unit of catalyst and per reactor is being used in the first stage of the synthesis plant in South Africa.

There are many ways in which one can bring catalyst and gas together and control the reaction, and it is a question of selecting the most promising lines and concentrating on those, and we are now going to concentrate on the liquid-phase system, which, to come back to your point, Sir, is essentially German in origin.

Lt.-Col. S. J. M. Auld: This has been a highly interesting communication. It has provided us with a great deal of fresh knowledge and appreciation of a new outlook on this problem. I hope that the authors in their turn have found it an opportunity, as it certainly has been for us, to provide a satisfactory and satisfying discussion. Thank you, Dr Hall and Mr Taylor, for your paper.

The vote of thanks was accorded with acclamation.

THE BARRACKPORE–WILSON OILFIELD OF TRINIDAD *

By G. E. HIGGINS (Associate Member)

GEOGRAPHICAL POSITION

THE field is located in the south-central area of Trinidad, in the County of Victoria, on the southern fringe of the so-called Naparima area (see Figs 1 and 2). It lies 7 miles south-east of San Fernando and is approximately 9 miles east of the Forest Reserve–Fyzabad producing area and 9 miles north-east of the Morne Diablo–Coora fields. It takes its name from the near-by village of Barrackpore and from Wilson, which was the name of the original lease owner of a 950-acre block lying to the south-west of Barrackpore.

PHYSIOGRAPHY

The field lies along the southern boundary of the Naparima peneplain belt of low, maturely-rounded physiography with a maximum relief of 150 ft above sea level, and the northern edge of the Oropouche alluvial and lagoonal flats. There is little or no relationship between physiography and structure. At the time of the discovery well, these flats were large swamp areas.

In the early 'twenties, the Government instituted an irrigation system and orderly drainage of the swamps began, while the reclaimed areas were extensively used, and still are, for the cultivation of rice and, to a lesser extent, sugar-cane. Nevertheless, the field extends into these flats, and particular care is exercised to avoid pollution troubles.

Drainage in the area is directed into the Oropouche River, which also receives water from streams draining from the southern ranges to the south. Flooding of the alluvial and lagoonal areas is frequent during times of heavy rain.

Since the developed area extends considerably into the lagoonal flats, road-making requires the building of embankments across the flat-lying areas to above flood level. Transport, until recently, was by small-gauge rail-car running along the top of the built-up embankment, while even now many of the wells are only accessible by rail-car. However, more extensive road-building has taken place in recent years.

CLIMATE

The climate is mild tropical with two seasons. The "dry" season extends from late December to May, and the "wet" season from June to December. The rainfall over the past ten years averaged 69·76 inches (range 53·84 to 88·19), of which an average of 75 per cent fell during the wet season. The daily rainfall during the wet season is variable, but as much as 4·7 inches has been recorded in one day. The prevailing winds are the trades, north-east during the dry season and south-east during the wet season. The average daily temperature lies between 80° and 88° F in the shade.

CAMP ESTABLISHMENT AND COMMUNICATIONS

Easy access to the field is available by means of second-class Government roads, from which Trinidad Leaseholds Ltd. (TLL) have constructed private branch roads.

FIG 1

The TLL now maintain a small establishment in the field area. Sixteen bungalows provide accommodation for staff. Field facilities comprise main office, stores, machine shop, garage, salvage shop, etc. Electric power, for field and domestic purposes, is supplied by diesel units. Water supply was originally obtained by damming, but is now supplemented by production from the freshwater sands logged in two abandoned wells in the Wilson area, and one well drilled specifically for these sands.

The oil produced travels by 3-inch and 4-inch lines

* MS received 2 November 1954.

to field storage tanks, where it is pumped at regular intervals through a 6-inch line from the field to Philippine, where it connects with the company's 8-inch and 10-inch trunk lines from Forest Reserve. These lines in turn connect through to the company's storage farm at Pointe-a-Pierre refinery. The gas production finds an outlet in power requirements for the field.

HISTORICAL AND SUMMARY NOTES ON DEVELOPMENT

Discovery and Development

The South Naparima Oilfields Co. Ltd. (SNO) drilled the first well in the area in June 1911. Oil was struck at a depth of 291 ft, and 14,568 brl were produced before the well was finally abandoned in 1942. The area received attention as a result of near-by oil and pitch seeps, whilst Cunningham-Craig had mapped an anticlinal structure there.

Subsequently, SNO drilled a further two wells before their holdings were transferred to Venezuelan Oilfields Exploration Company Ltd. (VOE), who drilled an additional six wells, and in 1914 transferred the SNO and VOE holdings to Trinidad Oil and Transport Co. Ltd. (TOT), who drilled a further three wells. In 1918–19, a local syndicate (Iere Oil Co. Ltd.) drilled three wells in the south-east edge of Barrackpore, of which only one was productive to the extent of 1465 brl, and the company has not operated since.

TLL took over the TOT holdings on 1 January 1918, and has developed the area ever since. Extension of the productive area proceeded south-westward., the eastern area in the so-called Moosrap and Gordon Plantations areas proving non-commercial. This area of early development is called the Old Barrackpore field.

In November 1935, TLL concluded an agreement with Trinidad Central Oilfields Ltd. (TCO) to develop the neighbouring Wilson block to the south-west. Wilson Well 1 was successfully completed in 1936 and development subsequently accelerated further south-westwards and included part of the TLL holdings in West Barrackpore. The contiguous properties to the Wilson area were under lease to United British Oilfields of Trinidad (UBOT), and development of their leases began in the early 'thirties. Since 1944 deeper drilling was carried out in the eastern area of Barrackpore and revealed deeper productive sands. Exploitation of these sands is now concentrated in that area and is proceeding by outsteps to the north-east.

The history of development in this field would be typical of the history of Trinidad's development as a whole, from shallow cable-tool drilling to the modern power rotary drill, with concomitant progress in depth and gradual overcoming of the major drilling difficulties. The rotary drill was brought in as early as 1916, but the weighting of muds was not introduced until the early 'thirties. The introduction of the Schlumberger log was a big advance, and the first well in the area to be logged was Wilson Well 1 in June 1936. Gamma-ray logging of accessible old wells was introduced in 1941. A recent development has been the introduction of oil-base mud for improvement in oil recovery and increased drilling efficiency, particularly through long shale sections in which drilling difficulties were encountered due to hydration and swelling of the surface active shales.

In common with other areas, fluctuating economic conditions affected the field development, and from time to time both optimistic and pessimistic views were held as to its future life.

Area of Development

The Barrackpore–Wilson field has been developed by TLL over an area of 4 miles, although the total extent, including the UBOT Penal development, is nearly 6 miles along strike. The width of development across strike is narrow and variable, but averages 1500 to 2000 ft. The approximate area so far developed by TLL amounts to 720 acres.

Wells Drilled and Production (as at 30 June 1951)

A total of 205 wells, including five deviated wells, have been drilled on the TLL and TCO properties. Of these, 103 had been abandoned at the end of June 1951. The depths of the wells range from 300 to 8000 ft, while the productive sands occur between 300 and 5500 ft, Miocene production extends to 3000 ft and deeper Oligocene production down to 5500 ft.

At the end of June 1951, 12,174,084 brl of oil had been produced, of which 468,034 brl were obtained prior to acquisition by TLL. Just over 2,000,000 brl have been obtained from the Oligocene sands developed since 1944–45.

Present Status and Future Development

Recently, the life of the field has been extended by virtue of the discovery of deeper (Oligocene) sands. This discovery was helped along by the successful results of UBOT in their deep drilling in Penal (1941). Although many geologists had previously recommended deeper drilling, their proposals had come to nought due to economic and drilling difficulties.

Exploitation of the Oligocene sands in the Wilson and West Barrackpore leases contiguous to UBOT proved restricted, and present exploitation for these sands is confined to the East Barrackpore area. Production, however, is variable and oil–water distributions are remarkable; in many instances it has become necessary to drill deviated holes from the original in order to find the oil-leg, deviations being up to more than 30°.

Future developments will be directed towards finding more Oligocene production, and further, to test the validity of the underthrust concept of structure in the hope of finding underthrust sands. This will involve deep drilling to at least 10,000 ft.

During 1948, a well was drilled on an apparent structural anomaly revealed by seismic surveys, about

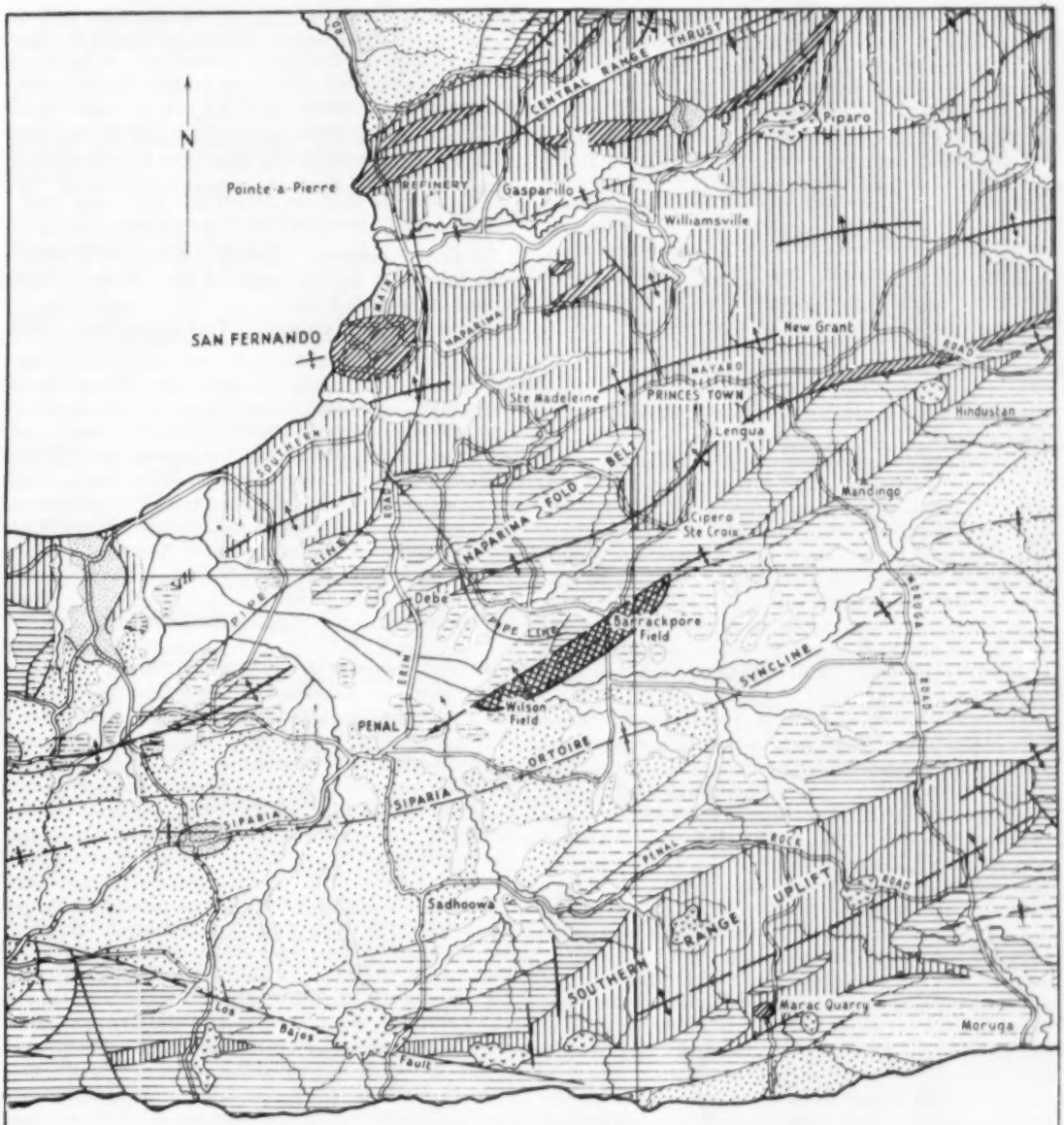

N
Pointe-a-Pierre
REFINERY
Gasparillo
CENTRAL RANGE THRUST
Piparo
Williamsville
SAN FERNANDO
MAIN
NAPARIMA
New Grant
NAVARO
PRINCES TOWN
Ste Madeleine
Lengua
Hindustan
Mandingo
SOUTHERN
MAIN
LINE
NAPARIMA FOLD BELT
Cipero
Ste Croix
Debe
PIPE LINE
Barrackpore Field
SYNCLINE
ERIN
Wilson Field
PENAL
ORTOIRE
SIPARIA
SIPARIA
PENAL ROCK UPLIFT
RANGE
SOUTHERN
Sadhoowa
Los Bajos
Fault
Marac Quarry
Moruga
MAP SHOWING
REGIONAL GEOLOGY & SETTING OF THE
BARRACKPORE – WILSON FIELD
TRINIDAD B.W.I.
0 1 2 3 4 5 MILES
LEGEND
Mudflow
Alluvium
Terraces (Pleistocene)
La Brea, Talparo, Morne L'Enfer (Pliocene)
Springvale, Forest. (Upper Miocene)
Brasso, Cruse, Lengua (Lower Miocene)
Cipero, Brasso, Karamat, Nariva (Oligocene)
Eocene, Paleocene, Cretaceous
Anticline
Syncline
Fault
Thrust Fault

FIG 2

¾ mile to the south of the Wilson field in the area of Clarke Road. Considerable difficulties were experienced during the drilling and completion of this well, but after re-drilling from 5000 ft, a successful completion was made at a depth of just over 10,000 ft. The significance of this discovery has yet to be assessed, and statistical data pertaining to this Clarke Road area are not included in this paper.

STRATIGRAPHY

The formations outcropping at the surface and in the wells range from recent to Middle Oligocene. The stratigraphic succession is outlined in Table I, whilst more detailed descriptions of the formations and other pertinent data are given below.

The succession given is sometimes arbitrary, for in common with other areas the general stratigraphic problems are by no means fully solved, particularly, for example, concerning the occurrence, relationships, and true thickness of the younger La Brea, Morne l'Enfer, and Forest formations.

Recent

This includes the alluvial and swamp deposits which fill the flats and river valleys south and west of Barrackpore. They are typical deposits and need no elaboration. They vary in thickness from a few feet to over 100 ft.

Pleistocene

Gravels and sandy terraces are found on the tops of hillocks throughout the area, the higher terraces appearing at 75± ft above present lagoon level.

Pliocene

La Brea Formation. This is not positively identified, but may include part of the Morne l'Enfer formation described below.

Morne l'Enfer Formation. This is generally easily recognized in well logs by the identification of freshwater sands. The sands outcrop to the south and west of Barrackpore, and in the Wilson area, where they form strong morphological features. The formation outcrops along the south flank of the uplift only, and is missing to the north. It is dominantly sand, well-bedded, fine to very fine-grained, with subordinate silt layers. The sands approach the ideal of sheet sands, and were believed to represent top-set deltaic beds in the past, although present opinion is not too clear on this matter.

The age of these freshwater sands is not certain,

TABLE I

Stratigraphic Summary: Barrackpore–Wilson Fields

Age		TLL zones—Biostratigraphic units	Formation	Beds	Lithology	Thickness	Producing horizon
HOLOCENE		—	River and swamp deposits—mud volcanoes, etc.		—	—	—
PLEISTOCENE		—	Terrace deposits?		—	—	—
PLIOCENE		—	La Brea	Freshwater sands (?)	Well-bedded sands with subordinate silts	?	No
		—	Morne l'Enfer	Freshwater sands	Well-bedded sands with subordinate silts	500 (+) ft	No
MIOCENE	Upper	*Nonion costiferum*	Forest	Wilson sands	Alternating sands, silts and shales; sands, lenticular and cross-bedded	3500 ft	Yes
		Guppyella miocenica		Lower Forest clay	Pure clay, shales with rare silts and sands	500 (±) ft	—
				Unconformable contact			
	Middle and Lower	*Discamminoides tobleri* *Alveovalvulina suteri*	Cruse	Lower Cruse clays, with occasional thin beds of calcareous (Lengua) clays	Pure clays, non-calcareous in general	400–4000 ft	No
				Cruse-Lengua beds, with vertical and lateral transition to Lengua and/or Lower Cruse.	Interbedded calcareous and non-calcareous clays		No
		Globorotalia menardii	Lengua		Calcareous clays		No
		Globorotalia mayeri			Calcareous clays with marly intercalations		
			Karamat	Unconformable contact(?) or transitional			
OLIGOCENE	Upper	*Globorotalia fohsi*	Cipero	Barrackpore–Penal Herrera Sands	Marls, calcareous and non-calcareous clays with "Pepper and Salt" sands	4000 (+) ft	Yes
		Globigerinatella insueta		Retrench Herrera Sands	Occasional radiolarian siltstones and rare reef limestones		
				St. Croix Limestone			
	Middle	*Globigerina dissimilis*	Nariva	(Unconformable contact)	Marls in general, with occasional calcareous and non-calcareous clays		No
	Lower	*Globigerina cf ciperoensis* *Globigerina cf apertura*	?				

and they may be partly Upper Miocene or comprise part of the Pliocene La Brea formation which is identified in South-west Trinidad. Leaf and other carbonaceous remains are noted, but lignites, such as those found in the La Brea formation, are not recorded, and apart from rare occurrences of pollen, *Buliminella* and *Nonion* species, the beds are palæontologically barren.

Not more than 500 ft of these beds are recorded in the wells, but southwards and westwards the thickness may attain at least 1500 ft.

Upper Miocene

Forest Formation. The subdivision of the Forest formation differs from that at the type locality of Forest reserve, where the division into an upper and lower member is usually well recognized. The upper Forest clay of the type locality is but poorly developed and seldom recognized palæontologically, or from electric logs. The lower Forest clay is, however, identifiable. Nevertheless, the sandy development of the Forest is concentrated above the lower Forest clay in this area, and it is convenient to subdivide the Forest formation into Wilson sands and lower Forest clay.

All the Miocene production of the area is obtained from the Wilson sands.

Wilson Sands. The name is derived from the Wilson area, where the formation is the major producing zone. The formation comprises an alternating series of sands, silts, and shales, with the sandy development greatest in the lower half of the formation. The sands are fine to very fine-grained, frequently laminated, and often poorly sorted with silty and clayey matrix. The sands may attain individual thicknesses in excess of 100 ft, but are extremely lenticular. They are typical of Trinidad's Miocene deposits, showing strong foreset bedding and irregularity in deposition. Bower has determined at least nine sand zones throughout the Wilson area, with an overall thickness of 3500 ft. The sand deposition shows a westward shift, and the successively younger sand zones appear further westwards to overlap either on the underlying Cruse formation or the clay–silt equivalents of the older sand zones. It is evident from Bower's study of the sand distribution that the sandy facies were laid down in a long (more than 4 miles) and narrow (1 mile) belt. The axial line of this elongate trough approximately parallels the crest of the Oligocene uplift, and it is inferred that this uplift limited the deposition of the sands to the north, since north of the uplift the whole Forest formation is in a dominantly argillaceous facies.

It should be noted that there is some uncertainty regarding the precise correlation, etc., between the Wilson and the Old Barrackpore field areas. The Old Barrackpore area is separated from the Wilson area by a zone of poor sand development, and exact correlation across this zone is highly conjectural in the absence of electric logs, etc. Of the 108 wells drilled in the Old Barrackpore field, only five were electrically logged, whilst a further sixteen were later surveyed by gamma-ray but without materially assisting the major correlation problems. There is little doubt, however, that the producing sands of the Old Barrackpore field are equivalent to the Wilson sands, and they are also comparable in thickness. The sands thin out to the north-east and are replaced almost entirely by silts and shales in the East Barrackpore area.

One feature of interest in these beds is the occurrence of slump zones. Beautiful examples of slumping have been observed at the surface and in cores. Frequently single cores show excessively steep dips, and it is quite a problem to state whether such dips are fault indications or slump phenomena without additional evidence.

The common diagnostic foraminiferal assemblages for the Wilson sands are composed of *Nonion costiferum* (Cushman), *Buliminella spp.*, *Cristellaria spp.*, and *Uvigerina spp.*

Lower Forest Clay. The Wilson sands directly overlie the lower Forest clay, and the contact appears normal but irregular in form. Only rare silts and thin sands occur within these beds, which are dominantly clays. The clays are dark grey to black, well-bedded, unctuous, non-calcareous, ferruginous, and weather to a red-brown colour. The lower Forest clay is rarely distinguishable from many other Miocene clays as far as lithological characteristics go, but is readily distinguished on a basis of its foraminiferal assemblage, which is characterized by a mixture of arenaceous and calcareous types, most of which have not yet been described.

The thickness of the formation increases away from the crestal area, where it is only a few hundred feet, to much greater thicknesses towards the south and west; the dominantly argillaceous character also becomes lost in this same direction and a more arenaceous development appears.

The base of the lower Forest clay marks a strong unconformity in the area, particularly in the crestal zone, where the clays rest directly on lower Cruse or older Miocene beds. To the south and west, upper and middle Cruse deposits appear. Although the well data show the contact to be unconformable, this unconformity has never been seen successfully in surface exposures, the contact being steeply dipping and probably complicated by fault and/or thrust phenomena.

The zonal marker of the lower Forest clay is the arenaceous foraminifera *Guppyella miocenica* (Cushman).

Middle and Lower Miocene

Cruse Formation. Within the area under review, upper and middle Cruse deposits are missing, being either non-deposited or eroded subsequent to the deposition of the lower Forest clay.

Lower Cruse Member. The lower Cruse is entirely

argillaceous. The clays are pure, unctuous, grey and black, non-calcareous, red weathering type. They are hydrophilic and plastic when wet. Landslips are frequent over the surface outcrops in these clays, and they are a drilling problem at depth.

Discamminoides tobleri Bronnimann and *Alveovalvulina suteri Bronnimann* are diagnostic species of the arenaceous lower cruse foraminiferal assemblages. In addition, we find *Cyclammina cancellata Brady* and *Bathysiphon sp. Cyclammina calcellata* has also been found commonly in the lower Forest and Cipero formations.

Lengua Formation. Towards the basal part of the lower Cruse, calcareous intercalations appear carrying a characteristic foraminiferal assemblage and the lower Cruse grades laterally and vertically into the Lengua beds, which are greenish grey, calcareous clays. The basal Lengua beds in turn show a strong affinity to the underlying Oligocene Cipero marls, and frequently show Oligocene foraminifera with evidence of reworking. In some cases the Oligo–Miocene contact appears transitional, but controversy still exists as to whether the contact does not represent true unconformable conditions rather than gradational.

The thickness of the lower Cruse–Lengua beds is variable but thin in the crestal and near-crestal areas (400 to 500 ft). However, a rapid thickening takes place southwards, especially in the Wilson area, where within the short horizontal distance of 3000 to 4000 ft the thickness appears to increase to at least 4000 ft. It is extremely doubtful whether this exceptional increase is due to repetition by faulting. It is now believed that the thickness can be explained by the interpretation of the Barrackpore structure as a fold with a strong southerly under-thrust limb which became a zone of collapse and infill during early Miocene time.

The Lengua is subdivided into two biostratigraphic zones, the diagnostic foraminifera being *Globorotalia menardii* (d'Orbigny) and *Globorotalia mayeri* (Cushman and Ellisor).

Oligocene

Cipero Formation and Herrera Sands. Oligocene formations outcrop to the north-east in Mandingo and beyond the present area of development. However, recent drilling in the East Barrackpore and West Barrackpore–Wilson areas has shown the Oligocene to be at comparatively shallow depth (5000 ft). The Oligocene is dominantly represented by the Cipero formation, which ranges from Upper to Lower Oligocene, although only Upper and Middle Oligocene have been recorded in the wells drilled to date. There is certain evidence indicating that the Cipero is also partly early Miocene, and future research will no doubt clarify this aspect. The Oligocene formations show variable lithologies from marls to pure non-calcareous clays with interbedded sandy horizons. Three separate facies types, which may interdigitate one with the other, are recognized. These types are:

(1) Marl, of open-sea, pelagic facies.

(2) Marine, clear-water, calcareous clays with occasional radiolarian siltstone (of the Retrench area) and rare reef limestone (of the St Croix area).

(3) Marine, turbid water, non-calcareous clays.

The Oligocene is exclusively subdivided on the occurrence of pelagic markers, and five main zones are recognized, corresponding to the *Globorotalia fohsi* (Cushman and Ellisor), *Globigerinatella insueta* (Cushman and Stainforth), *Globigerina dissimilis* (Cushman and Bermudez), *Globigerina ciperoensis Bolli*—this was formerly described as *Globigerina* cf *concinna Reuss*, and *Globigerina* cf *apertura Cushman*. The first two are considered to represent Upper Oligocene, whilst *Globigerina dissimilis* represents Middle Oligocene, and *Globigerina ciperoensis* and *Globigerina* cf *apertura* represent Lower Oligocene; this grouping is nevertheless an arbitrary one.

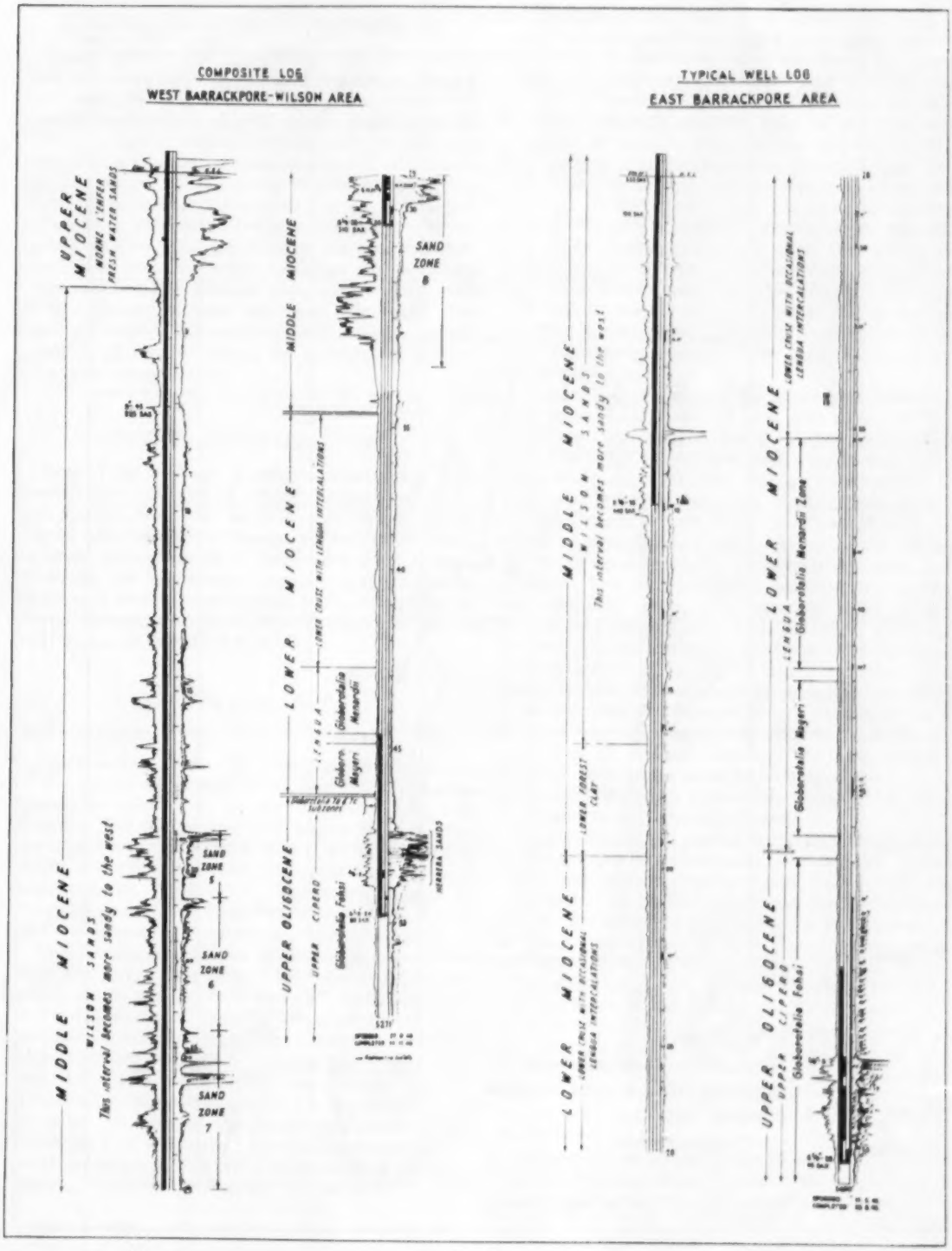
The Herrera sands have been found in association with the three facies mentioned above, and are characteristic so-called " pepper and salt " sands, due to the associated white quartz and black cherty, sometimes limestone, components. These sands are of the greywacke type, and show considerable variation in grain size, being conglomeratic in certain zones, but they are generally medium to fine-grained. Sorting is extremely variable. In the producing areas of north-east Barrackpore they attain a thickness of 250 ft or more, and may occur with rare interbedded silty and occasional gritty silt-clay bands. Throughout the whole area, a number of Herrera sand horizons are recognized, but there is still insufficient evidence to establish the correlation and inter-relationship of the various sands. The sands have been found principally in the *Globorotalia fohsi* zone, and extend along the length of the Penal–Barrackpore–Mandingo–tableland uplift, but with variable thickness from very thin to more than 1500 ft. They occur on the north flank of the uplift, but are non-commercial and in a generally finer-grained to silty facies. They are known to thin out farther north.

The so-called retrench Herrera sands of the *Globigerinatella insueta* zone have not been found in field wells, but they outcrop northwards.

The true thickness of the Oligocene formations in the Barrackpore area is not known, but so far at least 4000 ft have been penetrated on the north flank and at least 2500 ft in the south flank areas.

LOGGING CHARACTERISTICS OF FORMATIONS

Typical well logs of the Wilson–West Barrackpore and East Barrackpore areas are shown in Figs 3 (*a*) and (*b*) with identification of stratigraphic units. The Wilson–West Barrackpore log is a composite log, in order to show the freshwater and Oligocene (Herrera) sand horizons, which are not found in all wells.


COMPOSITE LOG
WEST BARRACKPORE-WILSON AREA
TYPICAL WELL LOG
EAST BARRACKPORE AREA
UPPER MIOCENE
MORNE L'ENFER FRESH WATER SANDS
MIDDLE MIOCENE
WILSON SANDS
This interval becomes more sandy to the west
SAND ZONE 5
SAND ZONE 6
SAND ZONE 7
SAND ZONE 8
LOWER MIOCENE
LOWER CROSS WITH OCCASIONAL LENGUA INTERCALATIONS
LENGUA
Globorotalia Mayeri
Globorotalia Menardii
UPPER OLIGOCENE
UPPER CIPERO
Globorotalia Fohsi
HERRERA SANDS
LOWER FOREST CLAY

(a) FIG 3 (b)

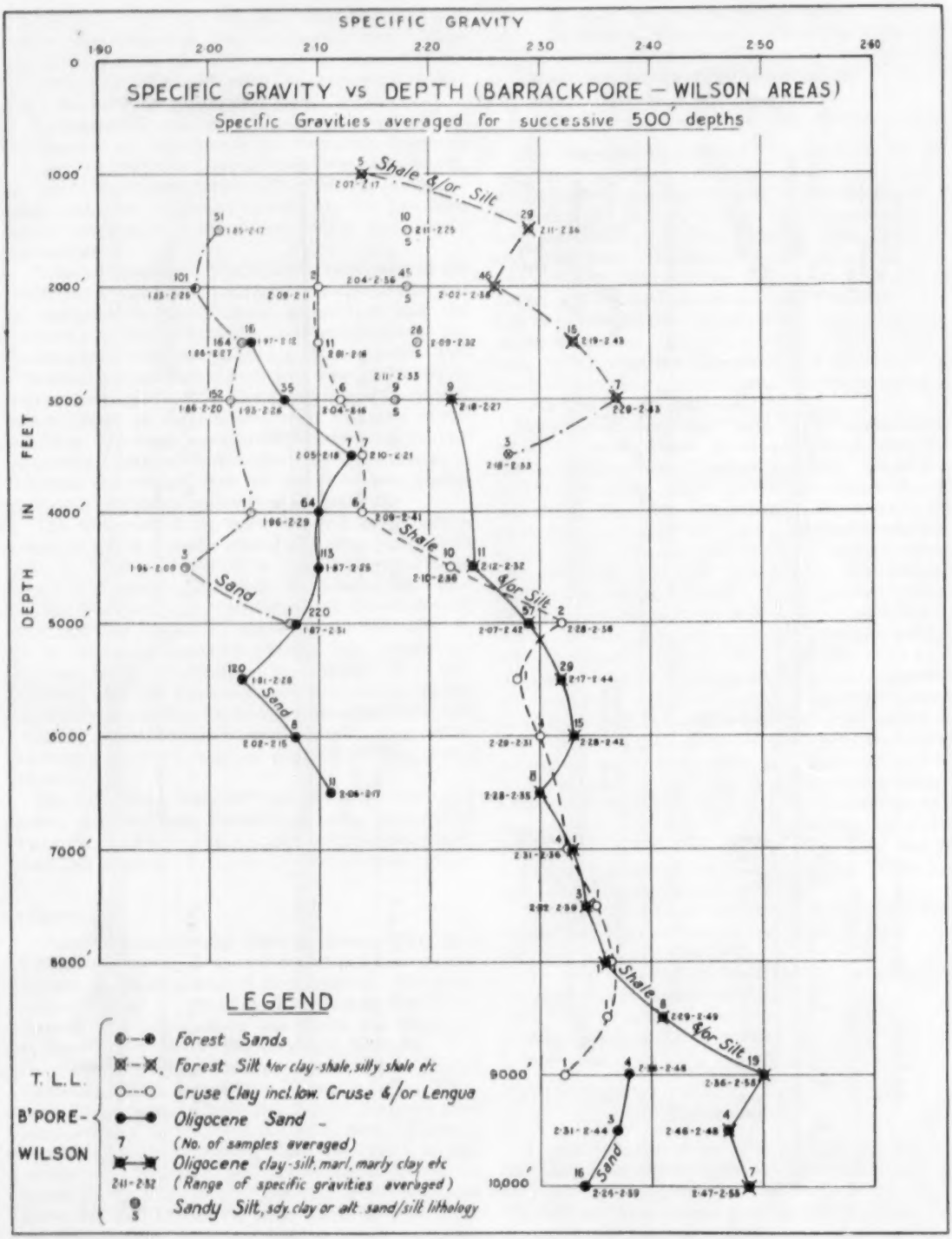

SPECIFIC GRAVITY
1·90
2·00
2·10
2·20
2·30
2·40
2·50
2·60
SPECIFIC GRAVITY vs DEPTH (BARRACKPORE – WILSON AREAS)
Specific Gravities averaged for successive 500′ depths
DEPTH IN FEET
1000′
2000′
3000′
4000′
5000′
6000′
7000′
8000′
9000′
10,000′
Shale &/or Silt
Sand
Shale
Sand
Shale &/or Silt
Sand
LEGEND
T. L. L.
B'PORE-
WILSON
Forest Sands
Forest Silt &/or clay-shale, silty shale etc
Cruse Clay incl. low. Cruse &/or Lengua
Oligocene Sand
(No. of samples averaged)
Oligocene clay-silt, marl, marly clay etc
(Range of specific gravities averaged)
Sandy Silt, sdy. clay or alt. sand/silt lithology

Fig. 4

SPECIFIC GRAVITY OF FORMATIONS

Specific gravity determinations on all core samples from field wells has been standard practice in the past few years, and Fig 4 shows the average results of these determinations plotted against depth, together with the range of values so determined. There is on the whole a gradual increase of density with depth, but the range of variations is considerable. The dominant factors are lithology and depth, with higher values being recorded in the silts and shales. Nevertheless, shallow, younger Miocene silts, for example, may show little difference in density from deeper Oligocene or older Miocene sediments, and in some wells a density reversal may occur at depth. The density column is not unambiguously normal, and the selection of suitable values for interpreting gravity data is no simple matter.

TEMPERATURE GRADIENT

Most of the available temperature data have been derived from wells drilled recently or producing wells, and static bottom hole temperatures are not always known with complete certainty, as the time involved to make measurements is limited by other factors. However, the temperature gradient appears normal in the area and averages approx 1° F per 75 ft depth. Recorded temperatures at 5000 ft depth are 155° to 163° F, and at 9900 ft 214° F.

GEOLOGICAL DATA

Regional Setting

The area forms part of that fold system lying south of the central range fold-thrust complex, which fold system is referred to as the Naparima fold belt (Fig 2). The thrusts and folds appear to have been developed by compressive forces acting from the NNW. and directed towards SSE., with the structural lines trending WSW–ENE. Folding diminishes in intensity southwards from the central range, but is still fairly strong in the southern range uplift area.

The Barrackpore–Wilson structure is part of the most southerly of the folds of the Naparima fold belt which can be mapped from surface observations, and is 8 miles south of the central range thrust. The field itself lies on the SW.-plunging nose of a structural high extending WSW.–ENE. from Penal in the south-west through Barrackpore, Mandingo, Hindustan, and Tableland to the north-east, over a length of approx 14 miles. Southward lies the Siparia syncline, the south flank of which rises to the fold system of the southern range uplift some 5 miles south of Barrackpore.

Surface Geology

The area (Fig 5) was first reconnoitred by Cunningham-Craig during his sojourn as Government geologist in Trinidad, 1905–7, when it was included as part of Craig's central anticline. Since that time, isolated surveys by Penny, Suter, *et al*, continued to resolve certain problems as they arose and Frischknecht made a preliminary survey in 1934, but the first comprehensive and detailed map of the area was compiled by Kugler, from various sources, in 1941. No up-to-date palæontological investigations had been made up to that time to separate the various members of the Miocene and Oligocene formations. In 1943 Rohr and Muhlemann thoroughly investigated the whole Barrackpore–Penal area throughout its length, and their final geological map, from which Fig 5 has been drawn, provided detailed coverage sufficient to resolve all problems as far as surface mapping is concerned. The only modification likely to be found as a result of future work will be on the interpretative side.

Surface mapping shows the outcropping Miocene beds to have an anticlinal form striking at WSW.–ENE. and plunging to the SW. The anticline is compressed and asymmetric, with a moderately steep north flank (about 45°) but with a very steep (80° to 90°) partly overturned south flank, which flattens rapidly towards the axis of the Siparia syncline to the south. Lower Miocene beds outcrop in the crestal region, which is a highly disturbed, squeezed, and shattered belt approaching a semi-diapyric nature with detached cores. The zone probably reflects longitudinal thrust faults, with the north flank up-thrusted; oil and gas shows are frequent in this zone. The south flank of the crestal region is marked by a line of oil and gas seepages with occasional mudflow. A second line of seepages occurs southward at a distance varying from 500 to 2000 ft, and appears indicative of normal strike or collapse faults—although such faults seem to be only minor ones as far as displacement is concerned. Some of the seeps are related to outcropping oil sands.

In addition to the normal strike and thrust faults mentioned, dip faults have also been mapped. These are of the tension type providing relief from the compressive forces that originated the Barrackpore uplift. The faults dip generally to the WSW. at about 60°, and some appear to have played an important part in concentrating accumulation.

The structure rises to the NE., and Oligocene marls and thin sands are exposed in the Mandingo area and beyond. The structure plunges to the SW., and in the Wilson area is hidden below the recent alluvial deposits. However, its extension is clearly revealed there as a result of the wells drilled, and the structure is seen to be much more gentle as far as the Miocene is concerned.

One feature of interest is the Digity Mud Volcano, which is located in an almost crestal position, in a slight saddle separating the Penal and Barrackpore areas. It is small, but typical of Trinidad's mud volcanoes, and has built itself up to about 30 ft above the level of the surrounding flats. The mudflow covers an area of about 5 acres, and salt water and

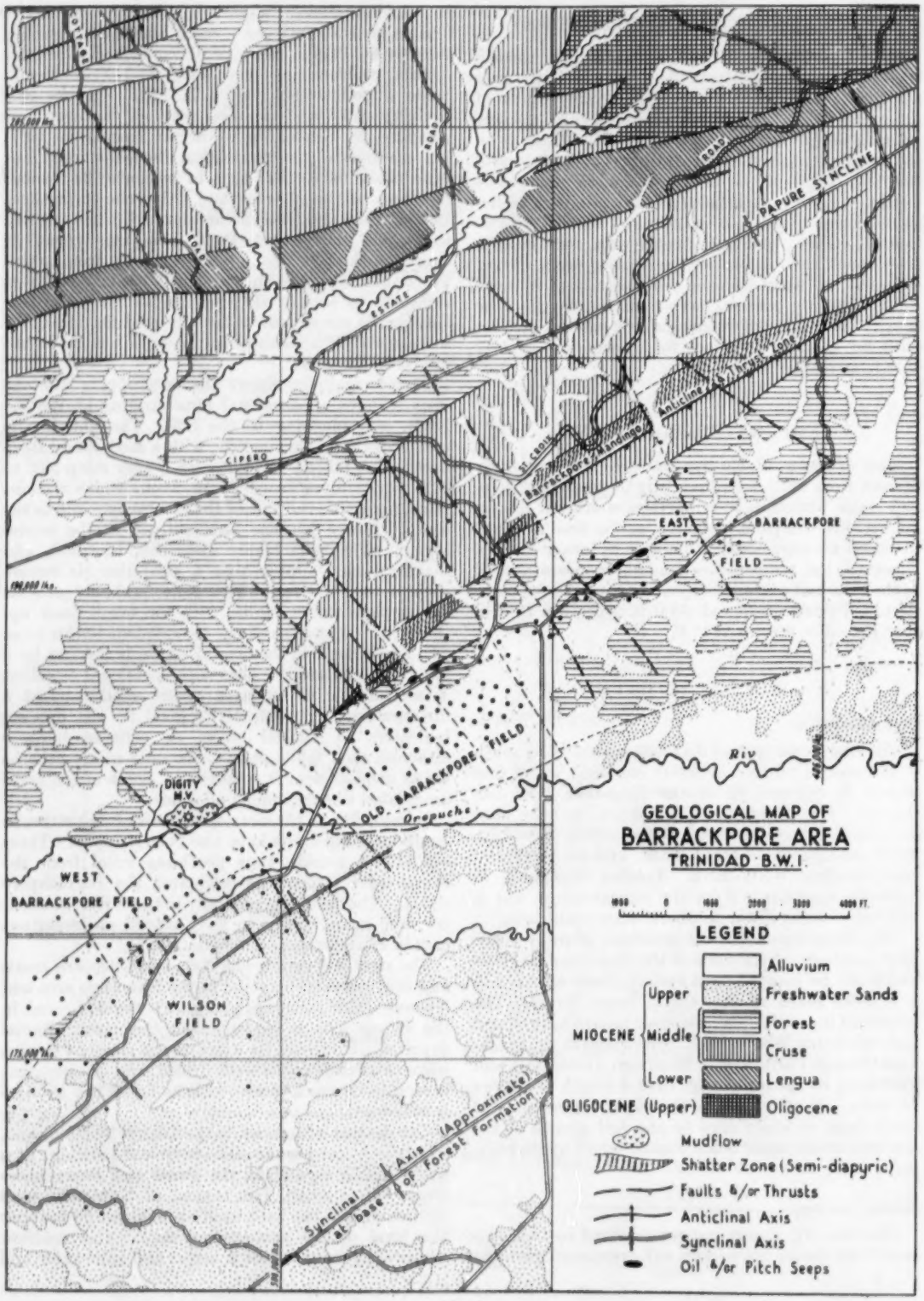

GUAYAD
ROAD
ROAD
ROAD
PAPURE SYNCLINE
ESTATE
CIPERO
Anticline
EAST
BARRACKPORE
FIELD
DIGITY M.V.
OLD BARRACKPORE FIELD
Oropuche
Riv
WEST
BARRACKPORE FIELD
WILSON
FIELD
Synclinal Axis (Approximate)
at base of Forest Formation
GEOLOGICAL MAP OF
BARRACKPORE AREA
TRINIDAD B.W.I.
LEGEND
Alluvium
Upper
Freshwater Sands
MIOCENE
Middle
Forest
Cruse
Lower
Lengua
OLIGOCENE (Upper)
Oligocene
Mudflow
Shatter Zone (Semi-diapyric)
Faults &/or Thrusts
Anticlinal Axis
Synclinal Axis
Oil &/or Pitch Seeps

FIG 5

gassified mud well up slowly through the apical vent of the main cone, whilst miscellaneous sluggish seeps are scattered around near-by. A very weak oil scum can be seen, and the phenomenon can be classified as a seep indicative of some subterranean reservoir below. The point of origin is most likely Upper Oligocene, for although the mudflow itself contains a foraminiferal assemblage of dominantly Lower Miocene (Lengua), Oligocene faunas are also reported, whilst the mudflow components include pieces of Upper Oligocene Herrera sands.

This separation is one of convenience from the point of view of geologic and other reasons as outlined below.

Old Barrackpore Field

This field was almost entirely drilled up before the advent of electric logging and the more modern techniques of core analysis etc. Information on the wells drilled in the area is too sketchy to permit a detailed appraisal of structure etc., but it is known with fair confidence that conditions are closely similar to those obtaining in the West Barrackpore–Wilson

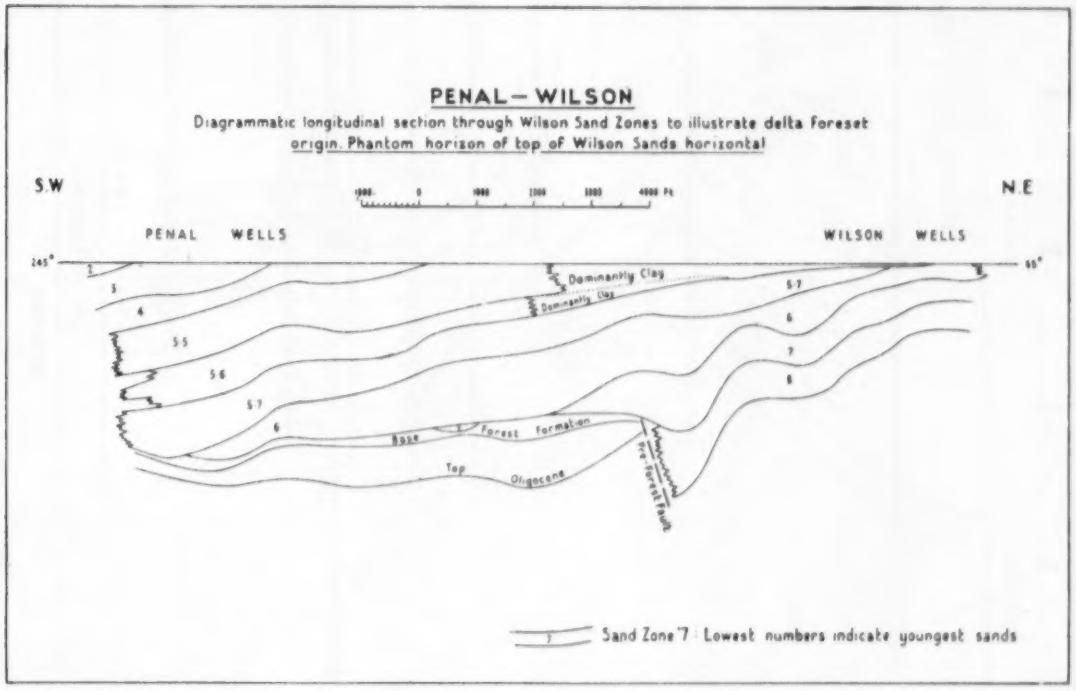


FIG 6

Structural Geology

The outlines of the structure as revealed by surface mapping have been described. Briefly, they show a compressed asymmetric anticlinal structure with a squeezed and shattered core, plunging to the WSW. The pitch of the structure appears to be of the order of 10°.

Production, however, is almost entirely confined to the south flank of the uplift, the exception being in the West Barrackpore area, where four wells have yielded Oligocene production from locations on the crest or slightly on the north flank of the uplift there.

It is preferable in discussing the area to view the production as derived from three field units:

(1) Old Barrackpore field.
(2) West Barrackpore and Wilson fields.
(3) East Barrackpore field.

M

area. Production is entirely from the Miocene Wilson sands, and accumulation is due to a combination of structural and stratigraphic factors. Strike faulting and the clay barrier of the crestal area have effectively acted as a seal limiting the northward migration of the oil, whilst in addition the sands shale out to the north and east, forming stratigraphic wedges.

West Barrackpore–Wilson Fields

Structurally, this area forms the south flank of the south-westerly plunging nose of the Barrackpore uplift. Of the seventy-five wells drilled in the area, five have been abandoned, and only four have produced from crestal or north flank locations, and in these cases the production is from Oligocene Herrera sands, with probably minor production available from Miocene sands. The discussion that follows is

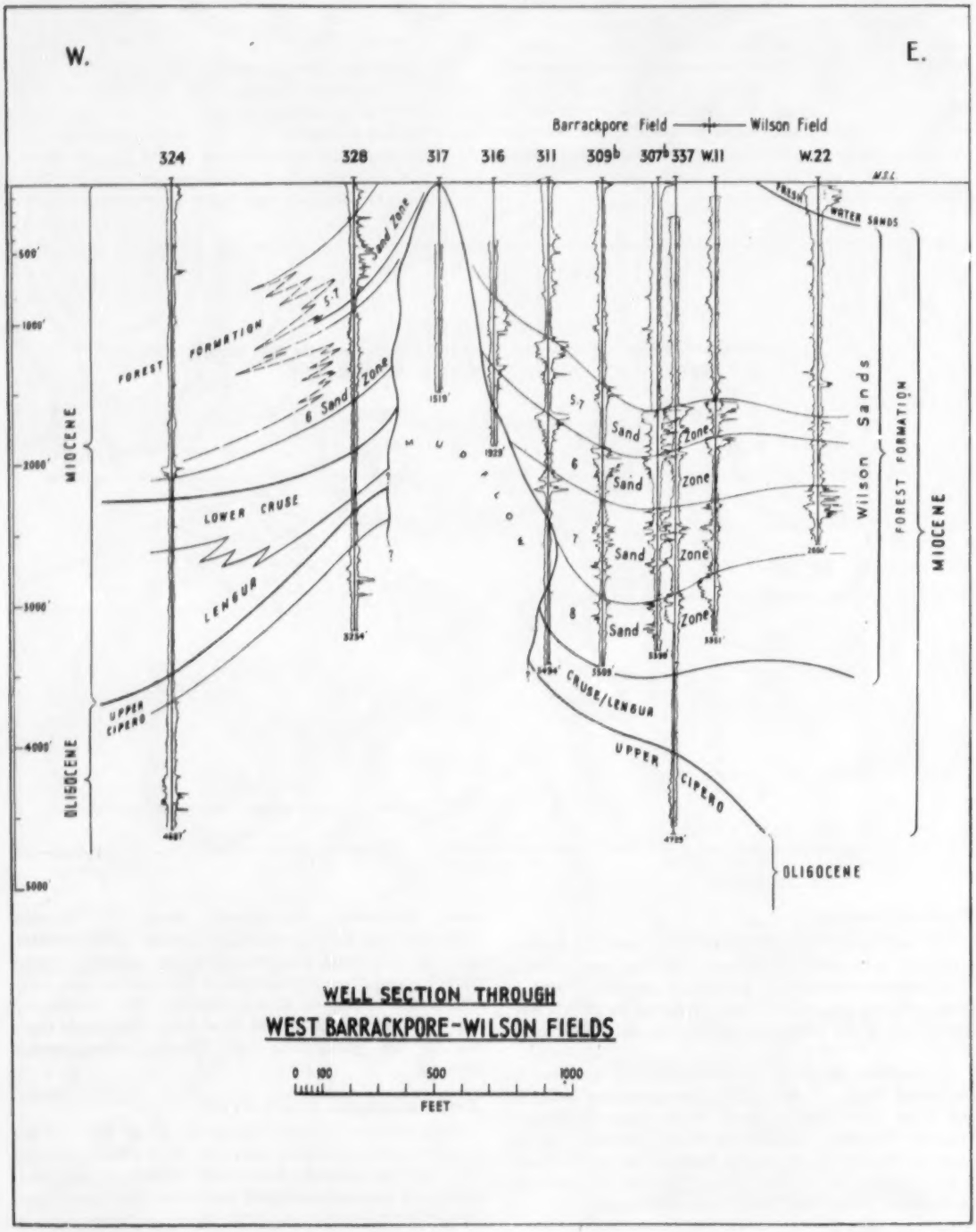

W.
E.
Barrackpore Field
Wilson Field
324
328
317
316
311
309b
307b
337
W.11
W.22
M.S.L.
FRESH WATER SANDS
500'
1000'
2000'
3000'
4000'
5000'
MIOCENE
OLIGOCENE
FOREST FORMATION
Sand Zone
5.7
6 Sand Zone
LOWER CRUSE
LENGUA
UPPER CIPERO
1519'
1929'
3234'
4487'
5.7
6
7
8
Sand
Zone
CRUSE/LENGUA
UPPER CIPERO
4725'
Wilson Sands
FOREST FORMATION
MIOCENE
OLIGOCENE
WELL SECTION THROUGH
WEST BARRACKPORE-WILSON FIELDS
0
100
500
1000
FEET

Fig 7

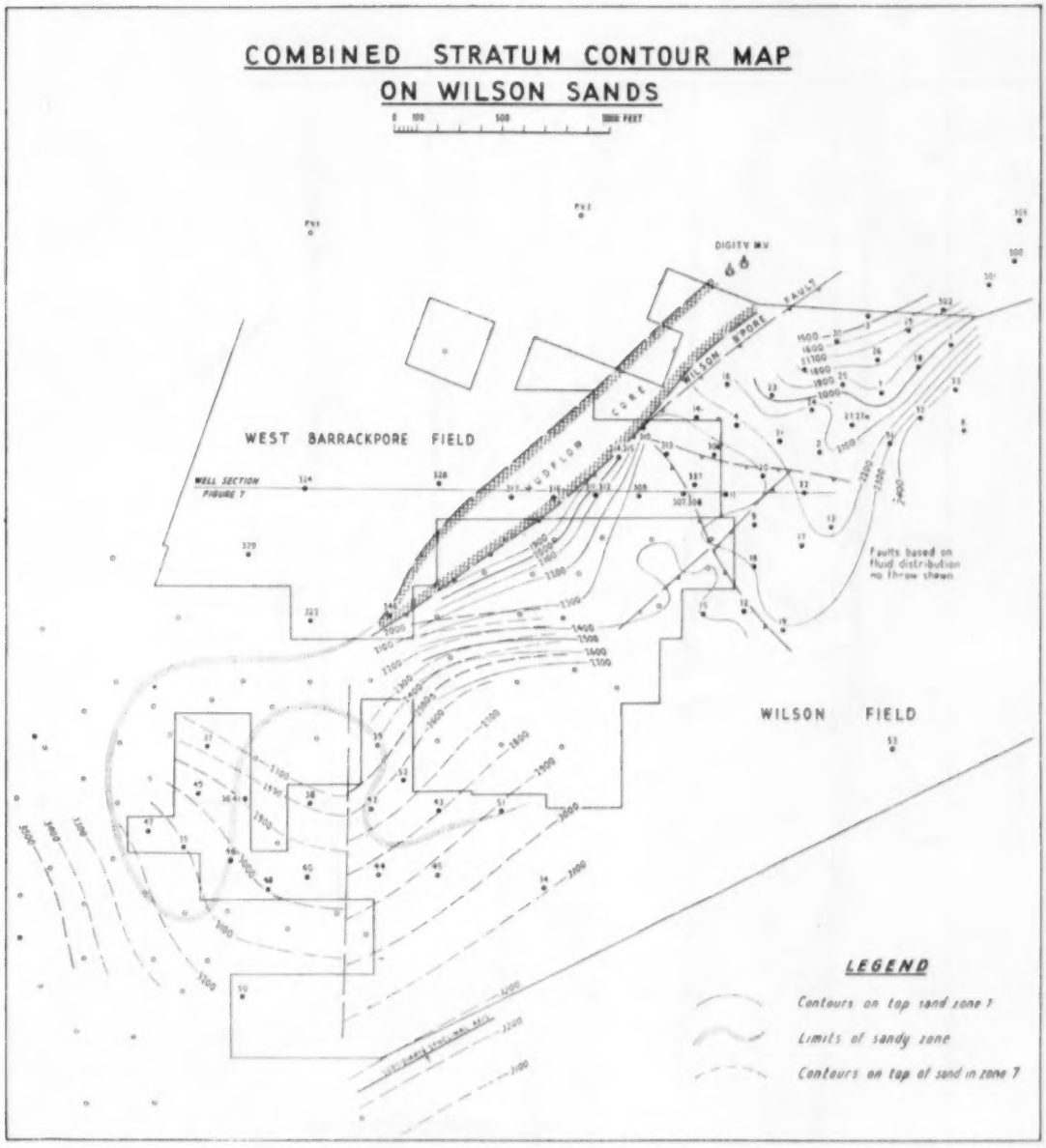

COMBINED STRATUM CONTOUR MAP
ON WILSON SANDS
0 100 500 5000 FEET
DIGITY M.V.
FAULT
WILSON PORE
CORE
MUDFLOW
WEST BARRACKPORE FIELD
WELL SECTION FIGURE 7
Faults based on fluid distribution no throw shown
WILSON FIELD
LEGEND
Contours on top sand zone 7
Limits of sandy zone
Contours on top of sand in zone 7

Fig. 8

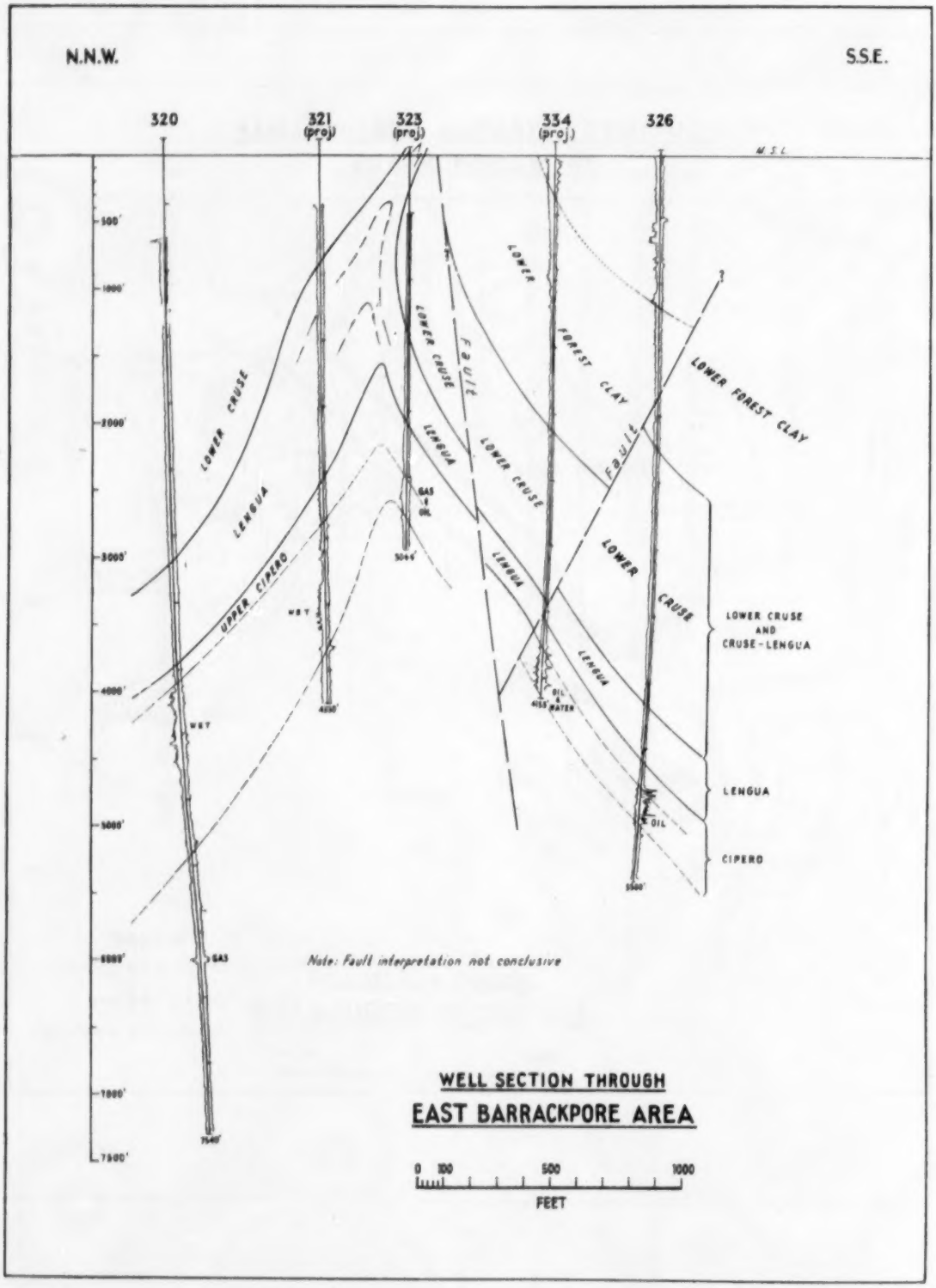

N.N.W.
S.S.E.
320
321 (proj)
323 (proj)
334 (proj)
326
M.S.L.
500'
1000'
2000'
3000'
4000'
5000'
6000'
7000'
LOWER CRUSE
LENGUA
UPPER CIPERO
LOWER CRUSE
LENGUA
Fault
LOWER FOREST CLAY
LOWER FOREST CLAY
Fault
LOWER CRUSE
LENGUA
LOWER CRUSE
LENGUA
LOWER CRUSE AND CRUSE-LENGUA
LENGUA
CIPERO
GAS
OIL
WET
WET
GAS
OIL & WATER
OIL
Note: Fault interpretation not conclusive
WELL SECTION THROUGH
EAST BARRACKPORE AREA
0 100 500 1000
FEET

Fig 9

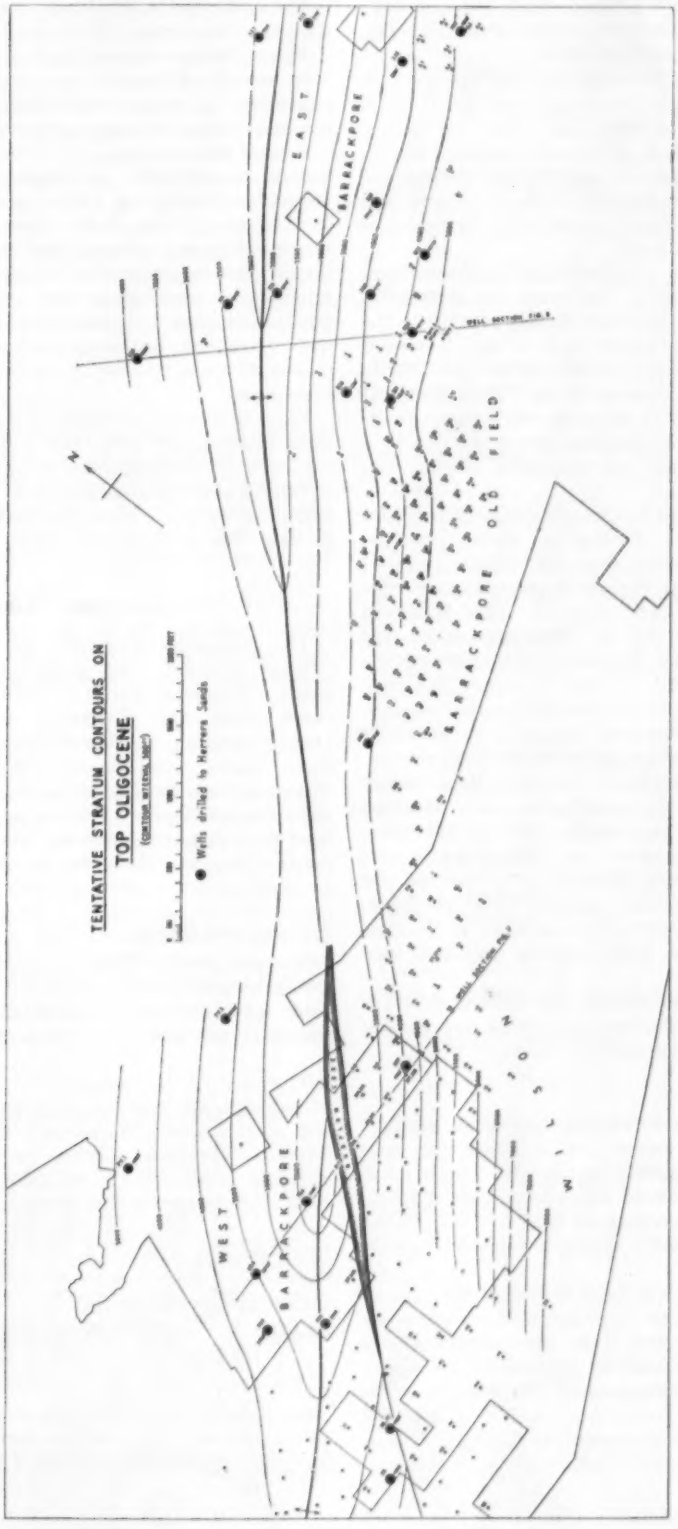

TENTATIVE STRATUM CONTOURS ON
TOP OLIGOCENE
Wells drilled to Herrera Sands
EAST
BARRACKPORE
BARRACKPORE OLD FIELD
WEST
BARRACKPORE
WILSON FIELD

FIG 10

concerned solely with the south flank Miocene wells which have produced over 90 per cent of the 5,470,000 brl of oil found in this area to date.

The area was drilled subsequent to the introduction of Schlumberger logging methods, and fairly good control exists, since all wells have been electrically surveyed. Nevertheless, it is to the credit of Bower, who carried out a detailed analysis of the well data in 1946, that a coherent picture of the structural and stratigraphic conditions within the West Barrackpore and Wilson fields now exists.

Accumulation is due to a combination of structural and stratigraphic factors: the sands dip steepest in the north, up to 30°, but they flatten rapidly to the south. The sands developed as local and lenticular bodies and are sharply truncated to the north, probably by the development of the Wilson–Barrackpore fault, considered to be a normal collapse fault. The crestal area of the plunging nose is defined by a thick mudflow core and its associated Digity Mud Volcano.

The field is remarkable for the extremely irregular oil–water distribution. Bower has shown that the deposition of the sands were of deltaic (foreset) character, being brought in and deposited in an overlapping series from east to west. This is shown diagrammatically in Fig 6. The deposition was probably controlled by long-shore currents, and the sands appear confined within a narrow elongate trough. This deposition yielded an ill-sorted mixture of sediments with pronounced changes in permeability and porosity, and was accompanied by local erosion, slumping, or infill, resulting in the irregular oil–water distribution. Although cross-faults are suspected between certain wells on a basis of fluid distributions, the characters of deposition and difficulties in correlation render it most difficult to determine the throw or locate the faults with any great accuracy. From the point of view of production, it is often necessary to produce wells entirely on their own merits.

A typical well section through the fields is shown in Fig 7 and a combined stratum contour map on the top of one of the Wilson sands is shown in Fig 8.

East Barrackpore

This area has been developed since 1944, and the productive Herrera sands are of Oligocene age. Miocene production in this part of the field is poor, due to the lack of sands, which, although well developed in the Old Barrackpore field, are thin in the north-east being almost completely replaced by silts and shales.

Twenty-one wells have been drilled in the area to date (this includes five deviated wells as separate wells) and of these, four have been outstandingly successful (with ultimates of 300,000 brl or more), two have expected ultimates of 150,000 brl, while production from the remainder could be considered the equivalent of one successful well. Eleven wells have been abandoned.

The core dips in the sands range from 10° to 60°, although the average dip of the flanks is about 45°.

The development and production of the area has been an enigma from the start, and still is, due to the anomalous oil–water distribution, which evidences a non-continuous, non-homogeneous reservoir.

Several theories have been developed to explain the anomalous oil–water relationships, and they cover ideas of tectonic, depositional, and grain-size distribution effects. None of the ideas can be said to be conclusive in any one well, and it is probable that all play a part in accounting for the irregularities. As a result, a cautious outstep programme has been pursued coupled with deviation of wells up-dip when the original hole has been found wet. This deviation (about 30°) has resulted in successful completions in three cases.

Fig 9 illustrates a typical well section through the East Barrackpore area, and Fig 10 shows stratum contours on top Oligocene, which is a simplified map for the whole Barrackpore and Wilson areas. The faults are omitted, since the data are insufficient to provide clear evidence that important faults exist.

RESERVOIRS

The producing zones are all sands, the Wilson (Upper Miocene) and Herrera (Upper Oligocene) sands. However, the environments under which these sands were deposited were very different; broadly speaking, the Forest–Cruse deposits were laid down under turbid-water conditions, whilst the Cipero deposits were laid down in clear water at about 1000 metres depth. The separate lithologies have been previously described in the section dealing with stratigraphy, but their general characters are further detailed here.

Lithology and Grain Size

Miocene Sands. These generally occur with two dominant grain sizes, very fine and fine, the former being most common. Laminated sediments are predominant, and sorting is very poor, though individual

TABLE II

Average Grain Size Analyses of Typical Miocene and Oligocene Sands

(In μ)

Age	Sediment, % sand	Coarse sand 1000–495	Medium sand 495–250	Fine sand 250–125	Very fine sand 125–63·5	Coarse silt 62·3–15·6	Fine silt 15·6–3·9	Clay less than 3·9
Miocene	80	—	2	36	45	11	5	1
	70–80	—	1	23	50	11	6	4
	50–70	—	—	14	44	25	8	9
Oligocene	80	1	14	38	34	6	3	4
	70–80	—	1	14	60	11	7	7
	50–70	—	—	7	63	18	10	12
	80	9	45	30	6	4	2	2
	70–80	—	25	53	14	4	2	2
	50–70	—	—	6	58	18	8	10

Note: 80 per cent sand comprises sediments in which the amount of very fine sand, fine sand, medium sand, and coarse sand exceeds 80 per cent of the total.

sands may show better sorting than Oligocene. Silt is a common constituent. The sands are generally non-calcareous; conglomerates are rare. A typical grain size analysis is shown in Table II.

Oligocene Sands. Sands of many sizes occur, including very fine, fine, and a large number of medium-grained sands. Laminated sediments are rare or absent, and sorting is poor. Conglomerates are common, and silt is relatively rare. Mixed sediments, such as gritty clays, are common. The sands are commonly calcareous, and sometimes cemented, due largely to calcite grains in the matrix of the rock. The calcite percentage may range from 5 to 40 per cent. In many respects it is not unlikely that these sands may be the products of turbidity flows.

Porosity and Permeabilities

Porosities of the Miocene sands are generally higher than the Oligocene, and range from 20 to 40 per cent, with an average of 30 per cent. Permeabilities are not fully known, as many of the oil sands are unconsolidated, while more consolidated layers have shown values of 120 to 600 millidarcies along the bedding.

Porosities of Oligocene sands range from 10 to 30 per cent, with the average at 25 per cent, whilst permeabilities are extremely variable, but average between 150 and 500 millidarcies along and across bedding.

Drive and Type of Reservoir

Drive is by gas-expansion, and the reservoir (Miocene and Oligocene sands) is of the so-called depletion type. There is not always a clear picture of edge-water conditions. Both Miocene and Oligocene wells show anomalous oil–water distribution.

As far as the Miocene is concerned, the oil–water distribution is separate for each individual sand zone, but the distribution is nevertheless in a sense controlled by down-dip position, with additional factors, such as changes in porosity, permeability, and probably faulting or slumping, also influencing the distribution locally.

In the instances of wells turning to water, the evidence is not always clear as to whether it is due to updip encroachment of water, or whether it is due to flooding from water sands immediately above or underlying the oil sands, as is particularly likely in the old wells, since canvas packers were often used for water shut-offs.

Tentative explanations for the irregular oil–water distribution in the Oligocene sands have been suggested on p. 140. More well evidence is required before the irregularities can be resolved.

Character of Oil and Water

Oil. Representative analyses of the characteristic Miocene and Oligocene oils are given in Tables III and IV, and Table V shows the average values of

TABLE III

Refinery Analysis—Typical Miocene Crude

(Well: Wilson 8)

Producing depth: 2508 to 2715 ft

Crude Oil

Sp. gr. at 60° F of sample	0·9472 (17·9° API)
Sp. gr. at 60° F dry oil	0·9470
Water (by distillation), % vol	0·3
Pour point, ° F fluid at	0
Viscosity, S.S.U. at 100° F	201
Total sulphur, wt %	0·67

Distillation of Crude Oil—Method R.L.E.–1

	% by vol
Light fractions (including loss)	8·2
Residue	91·8
	100·0
Distillation loss and error	0·2

Analysis of Products

	Light fractions	Residue
Sp. gr. at 60° F	0·7856	0·9590
ASTM distillation, ° C:		
I.B.P.	86	—
F.B.P.	210	—
10% at	120	—
30% at	133	—
50% at	143	—
90% at	179	—
Recovery, %	99·0	—
Residue, %	1·0	—
Loss, %	nil	—
Recovery at 100° C, %	3·0	—
Knock rating F-2	69·0	—
Flash point, (P.M.), ° F	—	196
Pour point, ° F	—	Fluid at 0
Viscosity, S.S.U. at 100° F	—	385

TABLE IV

Refinery Analysis—Typical Oligocene Crude

(Well: Barrackpore 334)

Producing depth: 3813 to 3836 ft, 4050 to 4065 ft

Crude Oil

Sp. gr. at 60° F of sample	0·8611 (32·8° API)
Sp. gr. at 60° F dry oil	0·8611 (32·8° API)
Water (by distillation), % vol	Trace
Pour point, ° F	0 or lower
Sulphur, total, wt %	0·37
Viscosity, S.S.U. at 100° F	42

Distillation of Crude Oil—Method R.L.E.–1

	% by vol
Light fractions (including loss)	34·6
Residue	65·4
	100·0
Distillation loss and error	0·1

Analysis of Products

	Light fractions	Residue
Sp. gr. at 60° F	0·7555	0·9095
ASTM distillation, ° C:		
I.B.P.	55	—
F.B.P.	198	—
10% at	88	—
30% at	111	—
50% at	129	—
90% at	170	—
Recovery, %	98·0	—
Residue, %	1·0	—
Loss, %	1·0	—
Recovery at 100° C, %	19·0	—
Octane No. F-2	52·0	—
Flash point (P.M.), ° F	—	200
Pour point, ° F	—	0 or lower
Viscosity, S.S.U. at 122° F	—	63

certain characters of the oils taken at random from a number of wells (twenty to thirty) in each field unit.

As a general classification, the Oligocene and Miocene oils may be determined as intermediate base oils, and from the point of view of refinery processing the oils are blended as one. Nevertheless, there are certain differences between the Miocene crudes of the whole area and the East Barrackpore Oligocene oils.

find if any relationship exists between produced water and connate water.

Table VI shows average results of data taken at random from the field wells, and it can be seen that the oilfield brines are very much alike for both Miocene and Oligocene waters of the West Barrackpore–Wilson and East Barrackpore areas. The Old Barrackpore data show somewhat higher total solids

TABLE V

Average Values of Certain Crude Oil Characteristics

Field unit	Producing zone	No. of wells averaged	Specific gravity of crude oil		Sulphur, % total weight		Pour point, °F		Viscosity at 100° F S.S.U.		Light fractions (L.F.), % up to 300° F		L.F : Octane No. ASTM	
			Av.	Range	Av.	Range	Av.	Range	Av.	Range	Av.	Range	Av.	Range
Old Barrackpore	Miocene	29	0·9312 (20·5° API)	0·8847–0·9537 (28·5°–17·0° API)	0·74	0·34–1·61	Below 0°	Below 0°	145	43–337	14·6	5·1–32·2	70·2	65·5–73·5
Wilson–West Barrackpore	Miocene	30	0·9446 (18·2° API)	0·8953–0·9775 (26·5°–13·3° API)	1·00	0·62–3·08	Below 0°	Below 0°	428	58–1767	11·0	2·1–27·7	67·4	63·0–73·0
East Barrackpore	Oligocene	20	0·8542 (34·2° API)	0·8188–0·9147 (41·5°–23·2° API)	0·36	0·14–1·14	Below 0°	Below 0°	45	33–98	29·8	14·5–42·0	58·6	50·4–69·0
Wilson–West Barrackpore	Oligocene	5	0·9190 (22·5° API)	0·9015–0·9402 (25·6°–19·0° API)	0·65	0·37–0·81	Below 0°	Below 0°	117	70–231	16·5	12·5–20·8	66·5	65·1–73·5

TABLE VI

Table Showing Average Results of Watch Analyses

	No. of samples	Age of sands	Parts per million								% Reacting values						Ratio—$Cl' : CO_3'' + HCO_3'$
			Total solids	$Na' + K$	Ca''	Mg''	Cl'	SO_4''	CO_3''	HCO_3'	$Na' + K$	Ca''	Mg''	$+HCO_3''$	SO_4''	Cl'	
Wilson–West Barrackpore	37	Miocene	20,793	7,761	27	34	8,918	16	348	4765	49·35	0·20	0·41	13·15	—	36·85	2·80
Old Barrackpore	26	Miocene	39,390	10,610	81	161	15,452	7	154	3511	47·16	0·40	1·35	6·42	—	44·63	7·0
East Barrackpore	12	Oligocene	19,912	7,610	60	39	9,023	43	146	4604	49·13	0·44	0·48	11·95	0·01	37·86	3·2

These differences can be seen in Table V and, briefly, are as follows:

(1) Oligocene oils are lighter than Miocene, and they yield a higher percentage of light fractions.

(2) Light fractions of the Oligocene oils are low octane (less than 64) as opposed to high octane for the light fractions of the Miocene.

(3) Oligocene crudes show a lower percentage total weight of sulphur than Miocene crudes.

The West Barrackpore and Wilson Oligocene crudes appear different from the East Barrackpore Oligocene crudes, and show closer alliance to the Miocene oils in general. The reason for this is not clear, and very little work has been done in regard to correlation of crude oil characters with geologic data.

Water. Routine analyses of any water produced in the wells is standard practice for the area, but it has not been found possible to use the data for correlation purposes, and although certain wells may show an unusual water, the source of the water is not always positively known. Analyses of connate water from cores are being carried out in certain wells to

and chlorides. The ratio of $Cl : (CO_3 + HCO_3)$ is also higher. The significance of the differences is not understood, particularly since the completion and production methods of the Old Barrackpore wells were often imperfect.

ORIGIN (SOURCE) OF OIL

There is little or no conclusive evidence regarding the source of the crude oils in this area. The occurrence of distinctive Miocene and Oligocene types suggests two sources, while in certain areas, *e.g.* the Wilson, there is some suggestion that the Miocene oil may have been derived from the underlying Oligocene or is perhaps contaminated with it. However, from general evidence and the close similarity with Miocene oils of other areas in Trinidad it is believed that the bulk of Miocene crude is indigenous to the formations in which it occurs. The Middle and Lower Miocene clays could have been the source rocks.

The source of the Oligocene oil is probably the associated marls and calcareous clays, which are highly foraminiferal. Nevertheless, there remains the possibility that pre-Miocene uplift and erosion could

have exposed the Oligocene sands, with consequent loss of original oil, although this exposure was not necessarily general for the whole area. In these cases, the Oligocene oil may have originated by downward migration from the Lower Miocene lengua–lower cruse clays or is a "second crop" Oligocene oil into the sands.

DRILLING

Method

All wells are now drilled exclusively by the rotary method, steam-powered rigs being used for all shallow and medium depth wells to 8000 ft, whilst diesel-powered rigs are used for 10,000-ft exploration tests. The earliest wells were drilled by the cable-tool method, the rotary being introduced in 1916 but not displacing the cable tool entirely until about 1920.

Major Difficulties

The more important difficulties met with in drilling wells in the early days were due to high gas pressures, heaving shale, and formation collapse. The introduction of barytes for weighting muds overcame the first, but heaving shales and formation collapse continue to be a problem, particularly where long (4000 ft or more) shale sections exist in the well. The shales are hydrophilic and in much of the area are steeply-dipping, highly slickensided, and fractured, which conditions almost certainly assist in the collapse of the formation. Oil-base mud has been recently introduced and is having marked success in eliminating drilling difficulties through these shale sections. The importance of good drilling mud to facilitate drilling of the wells is fully appreciated.

Rates of Drilling

The average time to complete wells to 3000 ft, after spudding in, is about four weeks, and for the medium depth wells to 5500 ft about seven weeks. This figure includes time for running casing etc.

CORING PRACTICE

In the early days of rotary drilling, coring was irregular but occasionally continuous through suspected sand zones, using the Spangler core barrel. Nevertheless, the cores did not receive the attention that is given to them nowadays.

In present practice, the geologists decide the coring programme for all wells drilled, and all suspected oil zones in undeveloped areas or where little information is available from previous wells, are cored continuously. Sidewall sampling (Howco, Thomas, or Schlumberger) may be carried out for palæontologic and/or oil–water analysis, but the recoveries by this method are not always satisfactory.

Cores from potential reservoir rocks are sealed on extraction and forwarded to the Forest Reserve field laboratory, where they are examined for porosity, permeability, saturation data, etc. Coring is done by conventional wireline barrel, and recoveries are variable but average 60 per cent.

WELL LOGGING

All wells are now electrically surveyed, and in the case of water-base wells the normal (AM = 9·5 inches), reverse (AM = 38·6 inches), and third curve (*i.e.* long normal, where AO = 17·94 ft) resistivity logs are run in addition to the normal S.P. curve. In the case of oil-base mud wells, scratcher logs have been tried with limited success, but are being replaced by the induction logging technique, supplemented with gamma-ray surveys.

Deviation surveys of wells are standard, and for dip surveys both Schlumberger E.M. and S.P. dipmeter surveys have been used as required. Temperature surveys are carried out in certain instances for location of top cement in cementing operations, while bottom-hole temperatures are now automatically incorporated with the Schlumberger electrical surveys of wells. Nevertheless, temperature surveys have been instituted on the more recent wells, and are recorded by the Humble temperature gauge.

DRILLING MUDS

The problem of mud for drilling programmes is considered an important one, and receives full attention at all times. The present-day practice has been built up in part from the trial and error methods of the early days. Exploitation wells follow a common programme for mud, whilst exploration wells show variations according to the conditions encountered. Sufficient material is kept on hand to combat unexpected conditions.

In the case of Miocene wells (3000 ft) the initial mud weights are approximately 80 to 85 lb/cu. ft. (11 lb/U.S.G.) and the muds are shale muds. This weight is stepped up to 92 to 96 lb/cu. ft. (12·5 lb/U.S.G.) for completion, by the addition of barytes. Certain Miocene sands appear to result in excessive wall-pack, and muds are treated according to conditions with various chemicals, such as quebracho and caustic soda, in order to provide the requisite gel strength, low water loss, viscosity, etc.

The Oligocene Herrera wells (5500 ft) commence with the same mud weights, but are stepped up gradually as the well progresses, so that on entry into the producing sands, weights are at 115 to 120 lb/cu. ft (15 to 16 lb/gal). The high mud weights are largely to combat caving. In one case, it was necessary to carry mud weights at 150 lb/cu. ft. (20 lb/gal) at 1000 ft depth, the well being finally drilled to 2500 ft with 135 lb (18 lb/gal) before being abandoned.

The introduction of oil-base muds raised new problems, which are being successfully overcome, and the mud weights used remain the same as for water-base muds.

CASING PROGRAMME

The present practice of drilling and completing wells utilizes two strings of casing. A conductor string of 13⅜, 10¾ or 9⅝ inches, generally the latter, and a completion string of 6⅝ inches. Provided no freshwater sands are logged by the well, the conductor string is set at about 100 ft for the shallow Miocene wells, and at about 900 ft in the deeper Oligocene wells. This string is cemented to the surface. The completion string is landed below the productive sands and sufficient cement placed at the back of the casing to provide for good water and fluid shut-off.

WELL SPACING

The spacing of Miocene wells has now been standardized at 400 ft, close spacing being desirable on account of the extreme irregularity and character of the producing sands. The Oligocene wells are at present developed on a 1200-ft spacing, since development is sensibly semi-exploratory. It remains to be seen whether interdrilling on this spacing for Oligocene production will be required at a later date.

PRODUCTION

The history of production practice in this area is one showing development from the uncontrolled gusher of the early days to the present standard of controlled (beaned) production, with due regard being paid to bottom hole pressures, gas–oil ratios (G.O.R.), rate of decline, and so on, in order to improve the efficiency of recovery.

Completion Methods

Present practices are normal, and, following the cementing of the oil string and fixing of surface fittings at the well head, the prospective sand zone or zones according to the log and individual thicknesses of the sand, are gun-perforated and the well is washed into production with oil. The use of water as a fluid to bring the well in has been discontinued in recent years because of the damaging effect that water may have on the reservoir rock. Wells are brought in through tubing and beaned as necessary according to standard production practice for control of G.O.R. etc.

The wells of the Old Barrackpore field were brought in before the use of tubing etc. became standard, and in these cases the completion of wells was controlled very much by the wells themselves. If they showed life, drilling continued as far as possible, and casing was then run with shop-perforated sections according to the anticipated oil sand depths. If necessary, the well was then bailed until clean oil showed, and produced on the casing with perhaps a ½- or 1-inch bean for control.

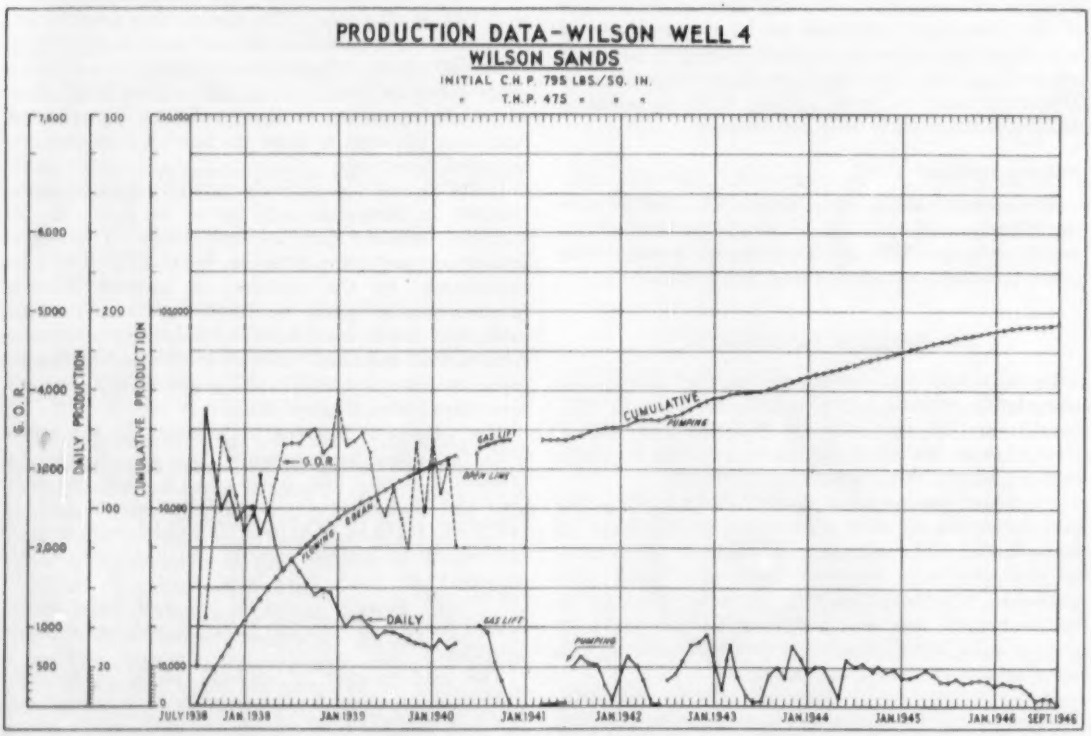


Fig 11

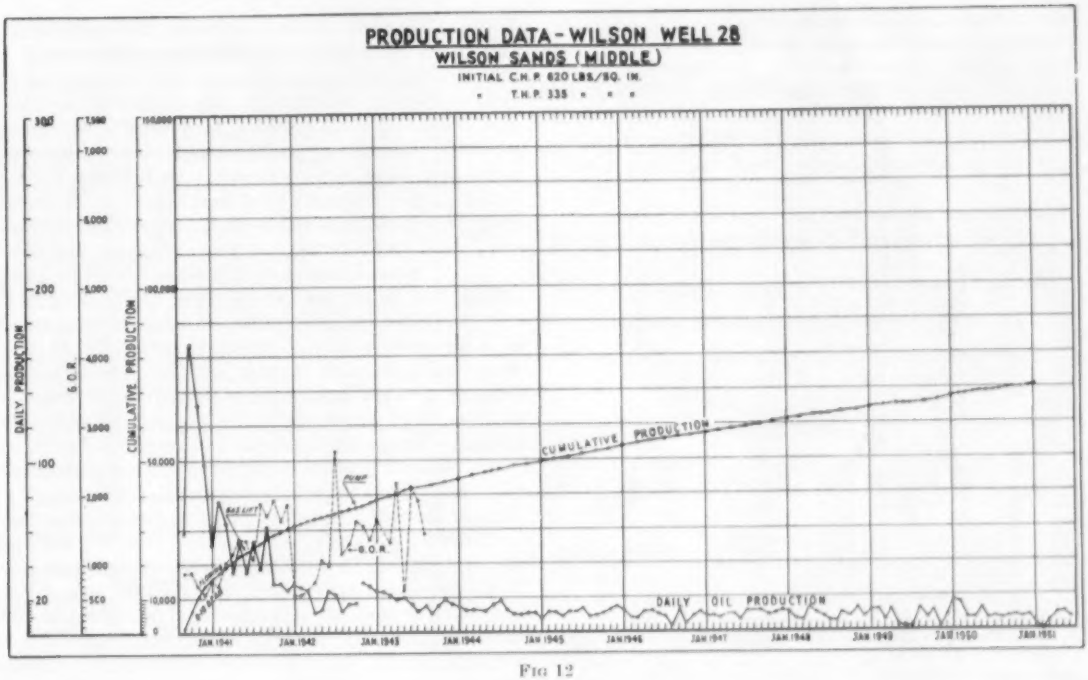

PRODUCTION DATA - WILSON WELL 28
WILSON SANDS (MIDDLE)
INITIAL C.H.P. 620 LBS./SQ. IN.
T.H.P. 335
DAILY PRODUCTION
G.O.R.
CUMULATIVE PRODUCTION
CUMULATIVE PRODUCTION
DAILY OIL PRODUCTION
G.O.R.
JAN. 1941
JAN. 1942
JAN. 1943
JAN. 1944
JAN. 1945
JAN. 1946
JAN. 1947
JAN. 1948
JAN. 1949
JAN. 1950
JAN. 1951

Fig 12

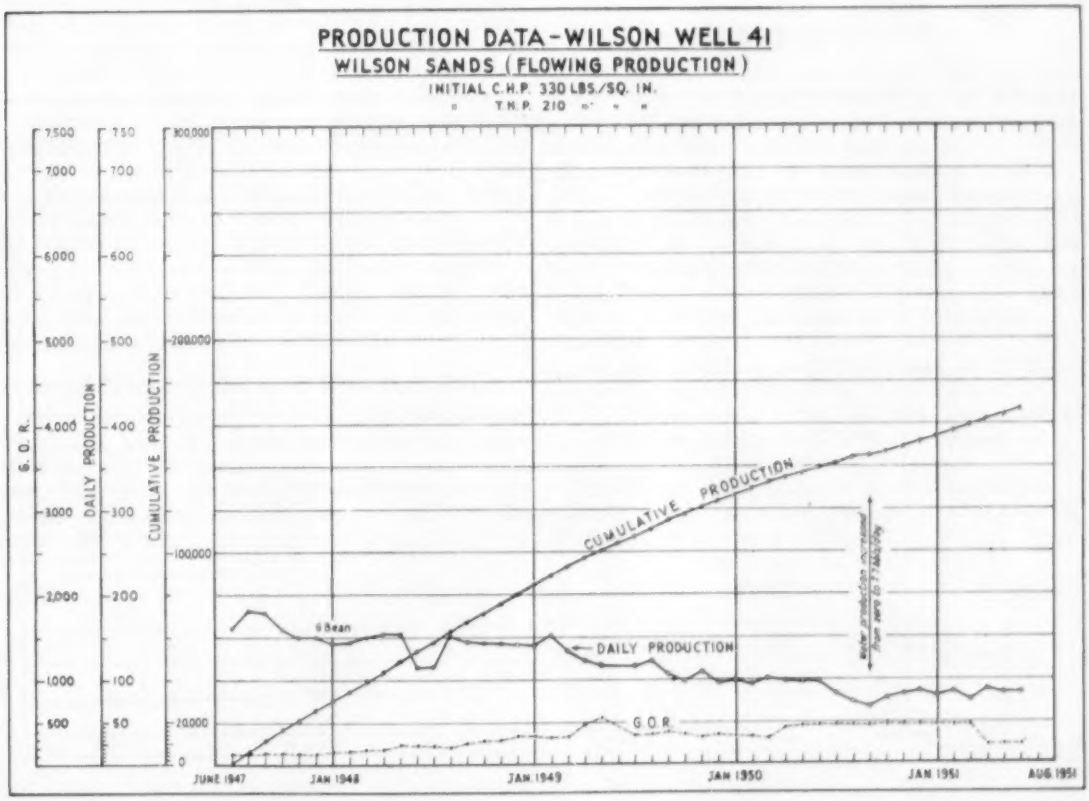

PRODUCTION DATA - WILSON WELL 41
WILSON SANDS (FLOWING PRODUCTION)
INITIAL C.H.P. 330 LBS./SQ. IN.
T.H.P. 210
G.O.R.
DAILY PRODUCTION
CUMULATIVE PRODUCTION
CUMULATIVE PRODUCTION
6 Bean
DAILY PRODUCTION
G.O.R.
JUNE 1947
JAN. 1948
JAN. 1949
JAN. 1950
JAN. 1951
AUG. 1951

Fig 13

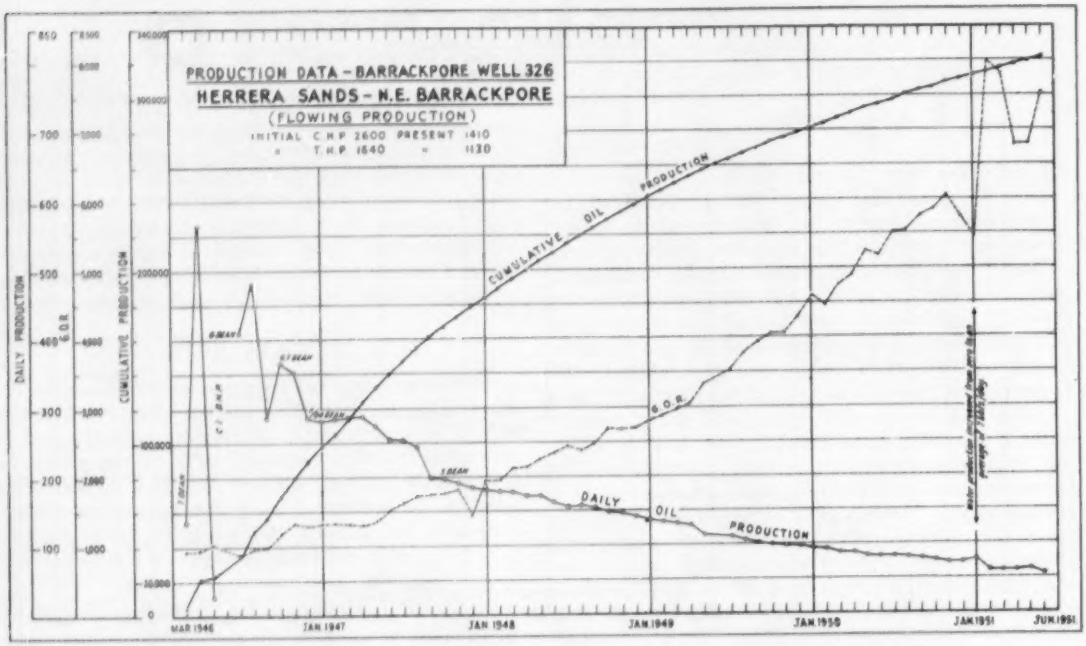

PRODUCTION DATA - BARRACKPORE WELL 326
HERRERA SANDS - N.E. BARRACKPORE
(FLOWING PRODUCTION)
INITIAL C.H.P. 2600 PRESENT 1410
" T.H.P. 1640 " 1130
DAILY PRODUCTION
G.O.R.
CUMULATIVE PRODUCTION
CUMULATIVE OIL PRODUCTION
G.O.R.
DAILY OIL PRODUCTION
MAR 1946
JAN 1947
JAN 1948
JAN 1949
JAN 1950
JAN 1951
JUN 1951

Fig 14

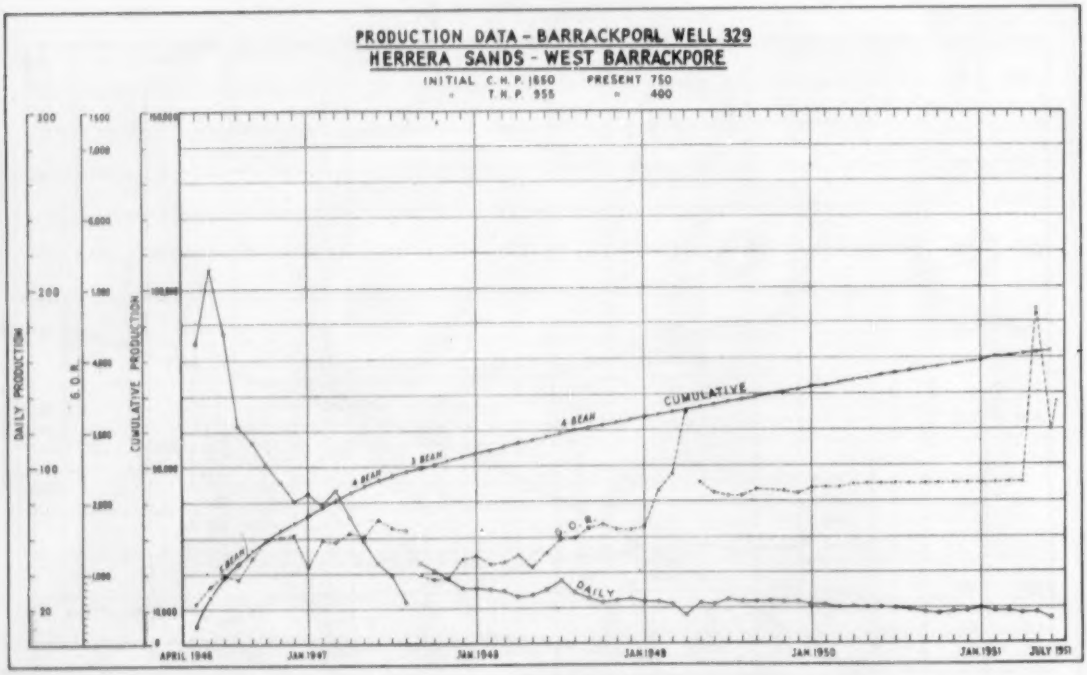

PRODUCTION DATA - BARRACKPORE WELL 329
HERRERA SANDS - WEST BARRACKPORE
INITIAL C.H.P. 1650 PRESENT 750
" T.H.P. 955 " 400
DAILY PRODUCTION
G.O.R.
CUMULATIVE PRODUCTION
CUMULATIVE
G.O.R.
DAILY
APRIL 1946
JAN 1947
JAN 1948
JAN 1949
JAN 1950
JAN 1951
JULY 1951

Fig 15

Production Data—Miocene Wells (Wilson–Barrackpore)

Early wells commenced life as flowing wells, with initial rates up to 200 b.d. on a $\frac{19}{32}$-inch bean with perforations over a zone of 100 ft approx. The more recent wells show rates varying from 50 to 100 b.d., depending on the zones open to production. Flowing life is controlled according to G.O.R., boundary agreements with competitors, production, etc. but declines fairly rapidly, varying from one to a few years, when wells are turned to pump. Gas lift has been tried, but pumping is generally more satisfactory.

Sand troubles occur in certain wells, and slotted liners have been found effective in some cases.

In the Old Barrackpore field, wells are pumped from a central unit, but the present practice in the more modern West Barrackpore–Wilson area is to have individual pumping units for the wells, resulting in an improved efficiency. Echometer surveys have become standard practice for determination of fluid levels.

Typical decline, annual, and cumulative production curves are shown in Figs 11, 12, and 13, for three wells in the Wilson area. The probable percentage recovery for the whole Wilson field is estimated at 8 per cent, though it is hoped that recoveries up to 20 per cent may be obtained in the case of individual sands. In many of the early wells, production was obtained from several sands at once, but nowadays it is considered preferable to exploit the well by perforating the lowest sand zones first, producing, and then when decline is such as to render further production uneconomic, the next successive higher zone is perforated and brought in. The highest percentage recovery yet recorded in one sand zone is 13 per cent.

Secondary recovery has not yet been seriously attempted in the area for obvious reasons, but may be tried in the future. An initial experiment in the Old Barrackpore field during 1934, utilizing air and gas for maintenance of pressure, was abandoned after six months on account of mudding troubles, probably caused by breakdown of the canvas packers originally used to shut off water sands. The lenticularity of the sands and non-uniformity of reservoirs will make secondary recovery a most difficult task.

Production Data—Herrera Wells (East Barrackpore)

All wells are brought in by gun-perforation and washing in with oil. Initial productions are higher than the average Miocene well, and flowing life is more prolonged. All producing wells at present are flowing. The initial production is variable, with the best at 500 to 600 b.d. on a $\frac{7}{32}$-inch bean, with 270 ft of perforations, but the average is between 300 and 450 b.d., on a $\frac{7}{32}$-inch bean with approximately 200-ft perforations. G.O.R.s are controlled and, although little b.h.p. data are yet available, they have been recorded in a few instances. The recorded pressure gradient averages 0·6 to 0·7 p.s.i/ft, with near crestal wells in the gas zone showing up to 0·85 p.s.i/ft.

Typical cumulative and annual decline curves of producing wells are shown in Figs 14 and 15.

Overall Production Data

Breakdown of total production is as follows:

Area	Producing horizons	No. of wells drilled	Date production first commenced	No. of wells producing at June 1951	Wells abandoned	Total production 1000 brl oil
Old Barrackpore	Miocene	109	1916	20	84	5139
Wilson–West Barrackpore	Miocene	75	1936	54	5	5030
Wilson–West Barrackpore	Oligocene	7*	1944	4	3	440
East Barrackpore	Oligocene	21†	1945	10	11	1575
Totals		212		88	103	12,174

* These wells are also included in Miocene total of wells drilled, since Miocene production is available from them. Three abandoned wells in this horizon are now producing from Miocene.
† Includes five deviated wells as separate wells.
Fourteen wells were idle at end of June 1951 awaiting repairs or abandonment.

ACKNOWLEDGMENTS

The information contained in this review has been taken from the reports available in the Trinidad Leaseholds Ltd. private files, and embodies the work of many company associates, past and present. It is impossible to record every name, although some have been mentioned in the review, and the writer is indebted to all for their help. He acknowledges especially the assistance of Drs Bronnimann and Bolli for stratigraphic and palæontologic data, the reports of Dr J. C. Griffiths for petrographic data, and T. H. Bower for much of the subsurface information.

The writer also expresses his thanks to Dr H. G. Kugler for his critical reading of the manuscript, to S. H. A. Mohammed for the original typing, and to R. A. Pallant and his staff for the preparation of the figures and maps accompanying the report.

Finally, the writer wishes to thank Trinidad Leaseholds Ltd. and Trinidad Central Oilfields Ltd. for permission to publish this review.

OBITUARIES

GEORGE MARTIN LEES

THE oil industry has thrown up its fair share of outstanding men in the last hundred years, and G. M. Lees must surely be numbered amongst them, although his place in that assembly is as distinctive as was the man himself. Of Scottish descent, he was brought up in Dublin until, in 1915, at the age of seventeen, he entered the Royal Military Academy at Woolwich, and in the following year saw active service with the Royal Artillery in France, where he was awarded the Military Cross. After joining the Royal Flying Corps in France, he was transferred to Egypt and later to Mesopotamia, where he earned the Distinguished Flying Cross in exploits behind the Turkish lines not far from the present Kirkuk oilfield. At the end of the war he was selected for duty as an Assistant Political Officer in Iraq under the administration of Sir Arnold Wilson. For some two years he then lived an adventurous life in the Halebja district of southern Kurdistan, where he gained an intimate knowledge of the way of life of those virile tribes and earned both their respect and affection. It was at this time, living amongst mountains which expose so much of their inner structure while hiding more, that Lees' interest was attracted to geology.

In 1920 he entered the Royal School of Mines with the intention of making geology his future career. In 1921, on the advice of Cunningham Craig, he accepted an appointment as Assistant Geologist with the Anglo-Persian Oil Company, rather than complete his course at the School of Mines, and was soon back in the Middle East, engaged in field work in Persia.

In 1923 he was selected by Professor de Böckh, then Geological Adviser to the A.P.O.C., to take part in a series of extensive reconnaissance surveys of South-west Persia, Iraq, and the Oman Peninsula which laid the foundations of our present knowledge of the geology of the Middle East. De Böckh realized Lees' potential as a geologist and a leader, and recommended that he should supplement his abbreviated academic training with a further period of study at the Vienna University, where Professor Kober's school of tectonic geology was then at its zenith. After taking his Ph.D. at Vienna with a doctorate thesis on the geology of the Oman mountains which is now a classic work on that region, Lees returned to reconnaissance work in Persia. He was appointed Chief Geologist of the Anglo-Persian Oil Co. in 1930.

During his twenty-three years as Chief Geologist, Lees directed and led the geological and geophysical work which enabled the Anglo-Iranian Oil Co. to develop four new oilfields in Persia and set the course of exploration so that further discoveries in that country may follow. His contribution to the discovery of great oil reserves in Kuwait, Qatar, and Iraq was also important, if not preponderant, and it can be said without question that the discovery of commercial oil in England, of which more than three-quarters of a million tons have already been produced, was primarily a result of his technical vision and energy.

Lees' appointment as Chief Geologist of the Anglo-Persian Oil Company coincided with the great slump of the early thirties, and his first task was therefore one of departmental retrenchment, an operation wholly repugnant to his constructive nature. Two years later followed the political crisis which led to substantial reductions in the size of the Persian Concession area, and to him fell the principal responsibility of selecting the territory to be surrendered. This involved the organization of a series of critical surveys and an appraisal of what the oil prospects of complex regions might be several decades ahead. To this grave task Lees brought the wide regional knowledge he had been acquiring since he joined the Company and a judgment which has stood the test of twenty years.

With the passing of the slump years, the Company again began to expand its exploration activities, and Lees gave much of his tremendous energy to the search for oil in the U.K. But for his persistent efforts in the face of considerable professional geological opinion and other obstacles the Nottinghamshire oilfields would not have been found in time to make their modest but important contribution to the defence of these islands during the second world war. During that war Lees' activities were manifold. His first concern was to ensure the greatest possible output of crude from the U.K., and an intensive campaign of exploratory drilling was put in hand. For part of the war he was attached to the Ministry of Fuel and Power to give of his extensive knowledge in the prosecution of the economic war. In this connexion he visited the Far East shortly before the Japanese invasion. Later, when the demand for oil supplies from the Persian Gulf again became pressing, Lees was moved to Persia.

For his last eight years with the Company, Lees was in the forefront of the increasingly competitive exploration of oil throughout the world and at the same time deeply involved with the rapid expansion of production in Persia. By this time he had visited every important oil producing area in the world and knew many of them intimately. Under his guidance the Company built up important exploration interests in Papua, Nigeria, East Africa, and elsewhere. In these years also his contributions to the science of petroleum geology reached their maturity, and he became the accepted leader of British oil geologists. His election as President of the Geological Society of London in 1948 and as a Fellow of the Royal Society in 1951 are evidence of the place be achieved in the scientific world. His achievements in the field of international petroleum geology were recognized by

the award of the Sidney Powers Memorial Medal in 1952 by the American Association of Petroleum Geologists. This honour, regarded as the highest professional distinction attainable by petroleum geologists in the U.S.A., has not been awarded to any other foreign geologist. He joined the Institution of Petroleum Technologists in 1924 and was elected to Fellowship when the name and constitution was changed in 1938.

Lees retired from his appointment as Chief Geologist of the Anglo-Iranian Oil Company in the latter part of 1953, but continued working as a consultant thereafter to the A.I.O.C. (subsequently B.P. Co.), the Burmah Oil Co., the Iraq Petroleum Co., and others.

When Lees embarked on his career as a petroleum geologist there was a tendency for oil men to eschew the academic. To be a geologist was almost synonymous for " practical oil men " with being something of an eccentric. If there had ever been a possibility of Lees becoming an eccentric, which in any case his good judgment would surely have precluded, his early experiences as soldier, airman, and political officer must have prevented it. It is characteristic of the man therefore that from his earliest days as a geologist he enthusiastically sought all that academic geology could offer. He became an active Fellow of the Geological Society, hitherto a fastness of pure science, and encouraged other oil geologists to follow him. His bringing together of academic and applied geology has done much to strengthen both sides of the science.

His principal interest lay in structural geology, a subject which requires the widest geological knowledge and the highest qualities of judgment and vision. His contributions to oil geology are best known through his many writings on the Middle East and the U.K., but his two Presidential Addresses to the Geological Society reveal more truly the depth and breadth of his geological thinking. From the end of 1952, when the illness set in which eventually proved fatal, until his death, Lees continued to contribute to the study of subjects of fundamental geological importance. He had hoped to contribute to a discussion on the Origin of the Earth arranged jointly between the Geological Society and the Royal Astronomical Society as a consequence of his last Presidential Address, but the meeting was held a week after his death.

For one who worked closely under Lees for considerable periods during the last twenty-five years it is difficult to compress into a few sentences an adequate appreciation of what he meant to his colleagues. He was so richly endowed with natural gifts, swift thinking, balanced judgment, physical energy, that at no time was his position of pre-eminence questioned. With such qualities some arrogance or intolerance might be pardoned, but there was no trace of either in Lees. Although a professional geologist, he gave " geology " as his only hobby in " Who's Who," but his zest for life in general was no less than that for his chosen subject. Humanity at large seemed for him to be a most intriguing part of a very interesting world. He looked at politics and international affairs, in both of which he was keenly interested, from the point of view of the individual involved rather than that of someone moving pieces on a board. As in geology, so in human matters, no accepted practice or principle was good enough until it had been probed and tested and found to withstand his penetrating analysis. Any form of humbug or insincerity was so contemptible to him that he rarely bothered to remark on it, but his forbearance and helpfulness in dealing with less gifted creatures was superlative. I have never met anyone who gave evidence of really disliking Lees. Indeed, he evoked almost universal affection as well as admiration, not only from those who had the opportunity to appreciate his greater achievements but also amongst the humbler folk with whom he came into contact. He seemed to have pruned away all the unnecessary and obstructing protuberances from his mind so that, like a master key, it would open all doors.

To travel with him through the mountain country of south-west Persia was an experience not to be forgotten. His keenness in working out the geological problems of the day, his quickness in finding the significant fossil or bitumen vein and in appreciating its implications, his visible enjoyment of a magnificent panorama or an unexpected gas seepage were stimulating enough, but in camp at night there were the arguments he provoked, enjoyed, and so adorned, whether on a geological topic or any other subject in the world. Often his enthusiasm during the day led so far afield that camp could not be reached by night and then bivouaced in some tribal village, he loved to enter into a semi-serious philosophical argument with his simple hosts.

That one who so enjoyed life and contributed so much to its colour as well as to human achievement should have died while still at the height of his powers is a tragedy we must mourn with his widow and son, both of whom meant so much to him.

P. T. C.

SIDNEY THOMAS MINCHIN

Born on 21 October 1902, Mr Minchin died in Farnborough Hospital on 11 January 1955, after an illness of nearly three months. After studying chemistry under the late Sir Robert Pickard, F.R.S., he graduated B.Sc., London, with first class honours in chemistry and was elected an associate of the Royal Institute of Chemistry. He joined the research staff of the Burmah Oil Company in 1923 and remained with that company for his whole career. In the early part of his service he specialized in the production and refining of kerosine, a commodity of particular importance in the Far East and subsequently devoted practically his whole attention to investigation into the manufacture and utilization of paraffin wax, in both of which fields by virtue of his able devotion

and initiative, he became an acknowledged expert, contributing a number of papers to the literature. For many years he was a member of sub-committees of the Institute of Petroleum dealing with the development of standard methods of testing and more recently in addition to being a member of the main Standardization Committee, he was chairman of the Petroleum Wax Sub-committee concerned with tests for the appraisal of these materials from the point of view of both producer and user. He was for many years a Fellow of the Institute.

Mr Minchin was of a very agreeable personality, and much thought of not only in his professional circle, but also amongst those interested in amateur acting. Author of several plays, he was in demand as a producer, and did much good work in connexion with the Beckenham Drama Festival and the Beckenham Children's Theatre. His advice and ever-ready help were sought and valued both by his business associates and outside parties alike. He will be greatly missed.

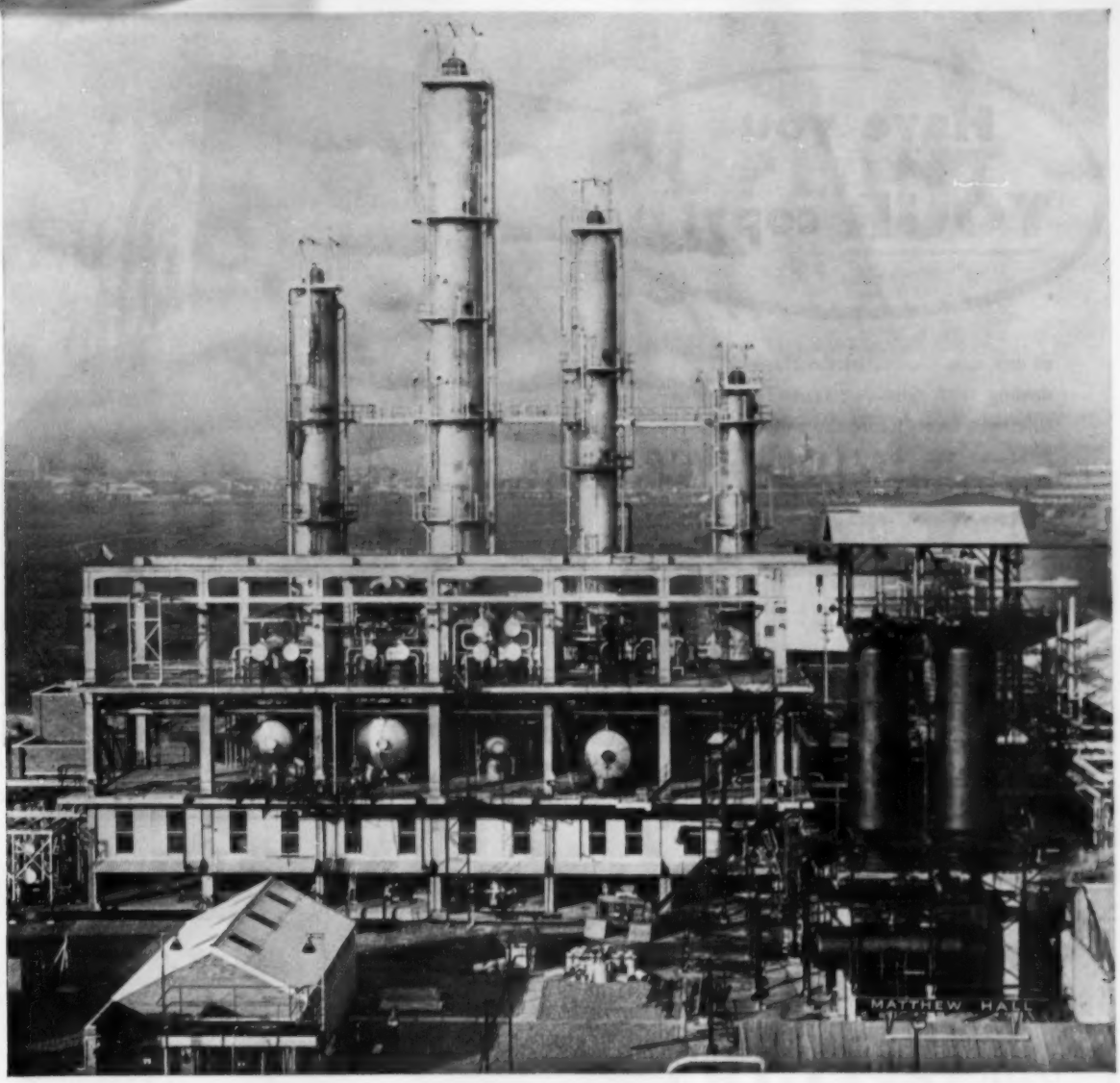
B. C. A.

DIESEL FUEL DILUENT—IP 22/53

The 1955 Edition of Standard Methods will be found not to include a method for estimation of diesel fuel diluent. First introduced in approximately its present form in 1944, a re-examination of the method was begun in 1948, because its precision had been found inadequate, and the method has been continuously on the agenda of the Miscellaneous Tests Panel since that time.

Initially attempts were made to improve the precision by superheating to a controlled temperature the steam introduced as a carrier for the volatile oil fractions, but the work was soon diverted to methods in which the test samples was partially distilled in the presence of an immiscible liquid of suitable boiling point, the vapour of which acted as carrier for the light petroleum fractions present. Thus brines of zinc and calcium chlorides, boiling at temperatures up to 150° C, were tried, and subsequently organic liquids in the glycol class. Initially promising results were obtained, but the problem became more complicated from about 1949 onwards, due to the strong trend towards the use of lubricants of considerably lower viscosity, which accordingly contained in themselves substantial proportions of material boiling at temperatures low enough to overlap the boiling range of the fuels used. A number of devices were examined, including the use of somewhat arbitrary methods for the interpretation of distillation curves, but no procedure was found acceptable for the assessment of fuel dilution without a considerable limitation of the scope of the method. Thus, it was concluded that no result upon a sample could be interpreted in terms of diesel fuel diluent without there also being available for examination samples of the actual unused lubricant and the fuel. Latterly the possible scope has been further narrowed because of the increasingly common use of fuels of high viscosity type, *i.e.* boiler fuel, in some types of engines.

Having regard to engineering progress in the diesel engine field, so that the occurrence of fuel dilution becomes a less serious problem, the need for a test method, and its utility even in its improved form from which greater precision would be obtainable, have been reviewed by Lubricants Sub-committee, and, taking into consideration the restrictions upon its scope as shown necessary by the studies completed, the decision taken to withdraw the test from "Standard Methods."



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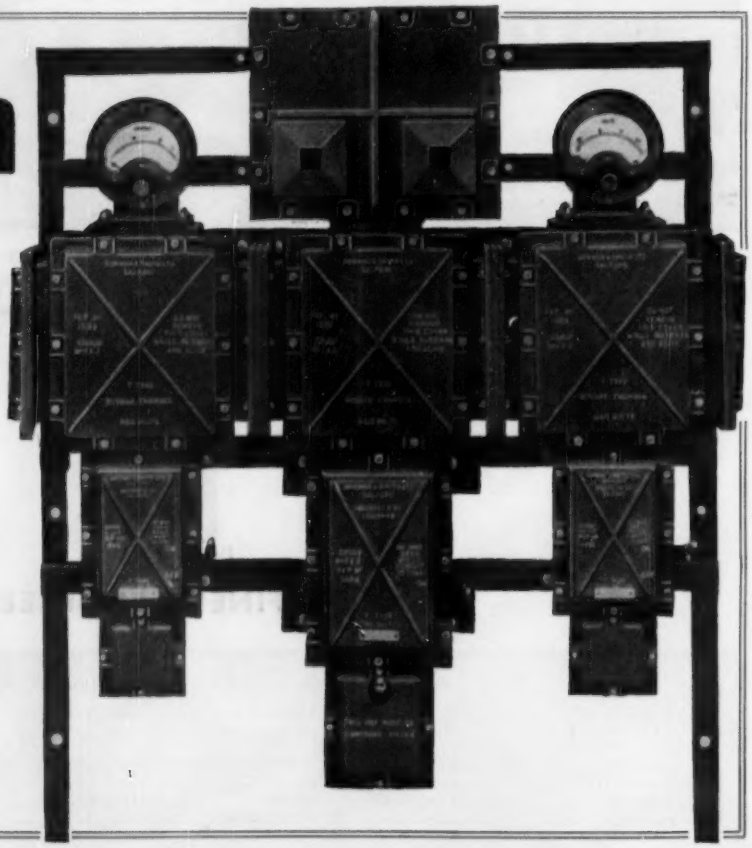

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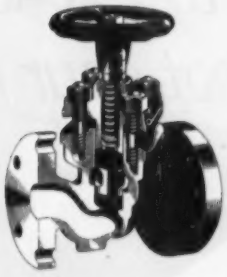

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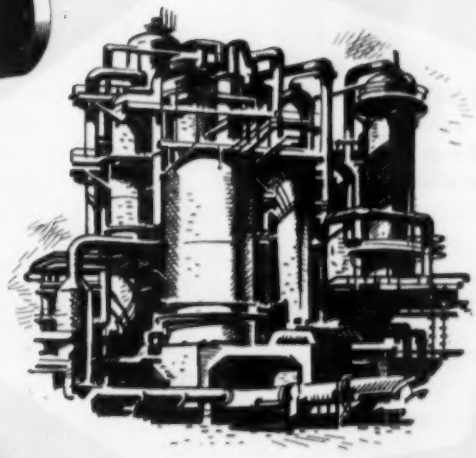
The Klinger Seatless Piston Valve operates on an entirely different principle from the usual type of seating valve. It consists basically of a finely ground piston, operated by the valve spindle and hand wheel and moving through two or three non-metallic resilient valve rings, capable of resisting steam, oil and most fluids at high temperatures and pressures, and separated by a ported lantern bush. It is made in a great variety of sizes and designs, to suit all purposes, and has the following advantages :—

No seat — consequently no regrinding.
Can be serviced in the line.
Unaffected by wiredrawing.
Unaffected by throttled use.



section

KLINGER REFLEX LEVEL GAUGES

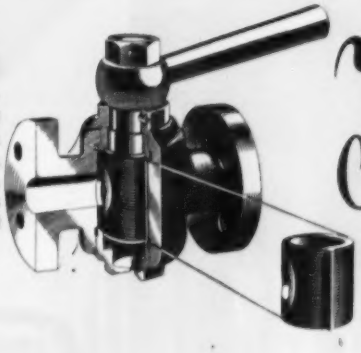
The "K" Type Reflex Level Gauge has the following outstanding advantages :—

Single row of tightening bolts.
Body free from distortion.
Glass easily removed.
Gauge body turnable, free to expand, and easily removed.

KLINGER SLEEVE-PACKED COCKS

In the last quarter of a century Klinger Sleeve-Packed Cocks have steadily replaced the old-fashioned asbestos-packed cocks with their limitations of working pressures, difficulty of repacking and tendency to jam. They are available in a wide range of designs and sizes for all purposes and pressures, and with the following advantages :—

Renewable "Klingerit" packing sleeve.
Retightening during use.
Parallel-ground non-jam plug.
Unobstructed straight-thru full-bore passage.



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an added "Range" for close tolerances.

15-RANGE
. . . for easy starting in "slim" holes, without "snubbing."

A size and type to EXACTLY meet your requirements

With the addition of the new "15-Range," Baker HINGE-LOK Casing Centralizers are now available for *all* combinations of casing sizes and hole sizes. From "slim" holes to large-diameter, oversize, irregular holes—or even where cavities are anticipated—there is a Baker HINGE-LOK Casing Centralizer to start without "snubbing" and to provide a *positive, uniform annular cementing space around the casing.* • Any Baker representative or office is ready to help you get "first-time" cementing results with Baker HINGE-LOK Casing Centralizers. To *end* right, *start* right—call BAKER.

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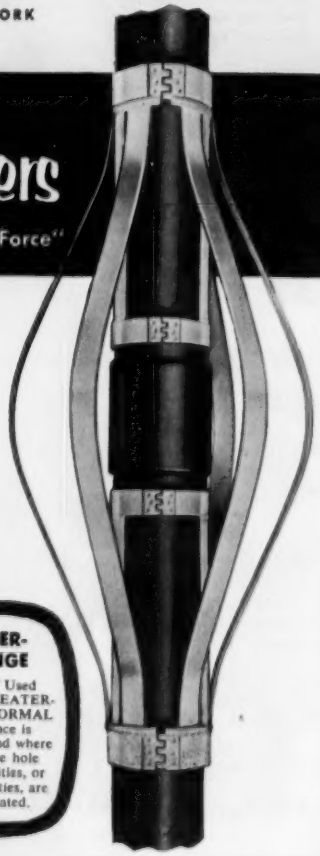
BAKER HINGE-LOK Casing Centralizers

always provide "Maximum Centering Force"

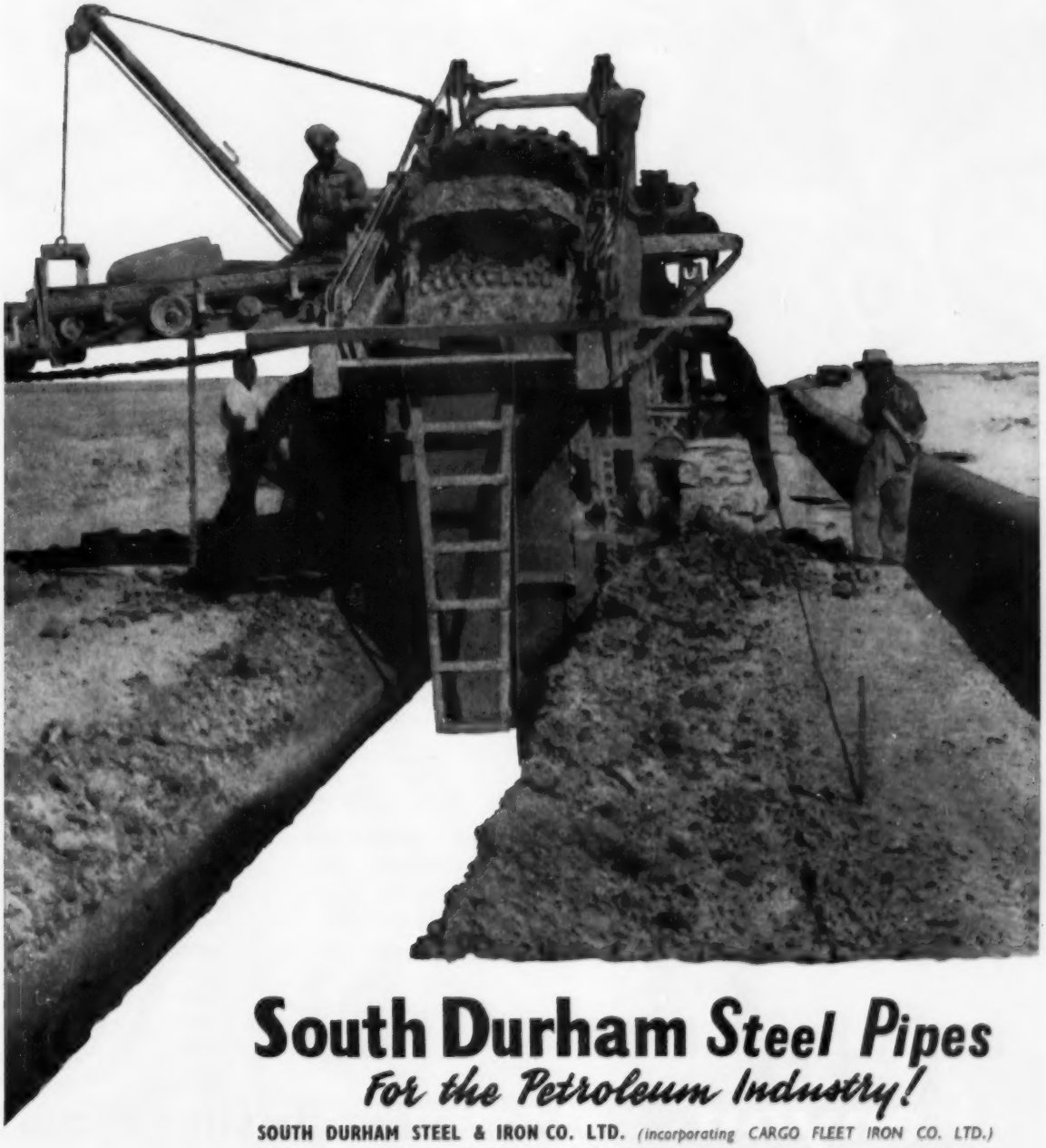
20-RANGE
Used where LESS-THAN-NORMAL clearance is present between the casing ON which centralizers are run and the casing THROUGH which they are run.

25-RANGE
Used where NORMAL clearance is present, and no serious problem of over-size hole is anticipated.



SUPER-RANGE
"H-50" Used where GREATER-THAN-NORMAL clearance is present and where extensive hole irregularities, or even cavities, are anticipated.

For any type of country!



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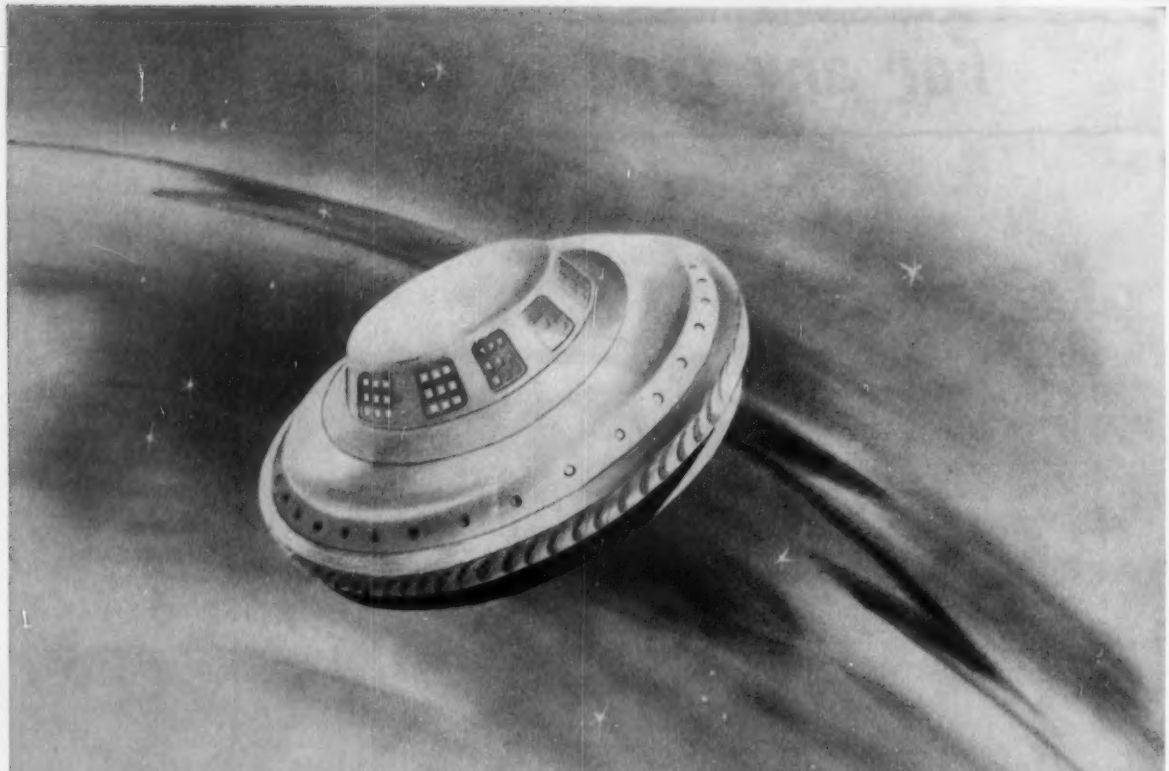

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Flying saucers? Each mystery that science resolves points to a greater mystery beyond. Indeed, there is no reason why we should believe earth to be the only inhabited planet in the universe.
But one thing is certain, if space travellers do exist the power used to drive their space craft is something quite phenomenal.
Coal and oil is still *our* greatest source of power — and Kenyon Planned Heat Insulation the greatest aid to its economy.

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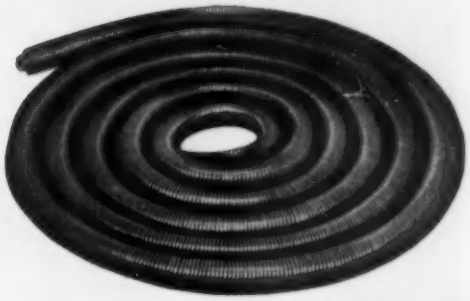
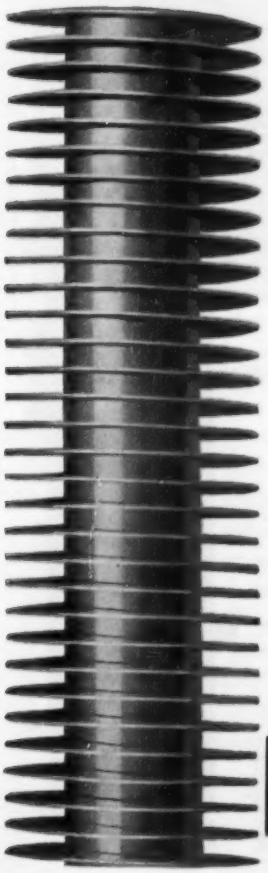
KH 174

'INTEGRON' Integral finned tubing





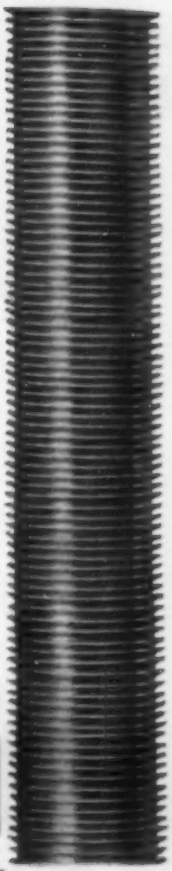
'INTEGRON' is a finned tube in which the fins are cold-formed from, and integral with, the parent tube. Made in copper, aluminium and certain alloys of these metals, 'Integron' tubing offers the advantages of strength, compactness, corrosion resistance and good heat transfer properties.

'Integron' is supplied in two forms—high-fin and low-fin —and in a wide range of fin spacings and dimensions to meet various heat exchange requirements.

The fins are not affected by vibration, shock or sudden temperature change and the continuous smooth surface reduces possibility of dust accumulation. 'Integron' tubing is easily manipulated—can be bent to close radii.

For radiators, convection heaters, skirting-board heaters, feed heaters, hydrogen and air coolers for electricity generators, oil coolers, transformer and rectifier oil or water coolers, compressed-air coolers, diesel-oil coolers, motor-car heaters, chemical coolers, 'Arcton' coolers, evaporators and condensers, humidifiers, dehumidifiers—and many other heat exchange applications.

Literature available on request.



'INTEGRON' finned tubing

IMPERIAL CHEMICAL INDUSTRIES LIMITED, LONDON, S.W.1

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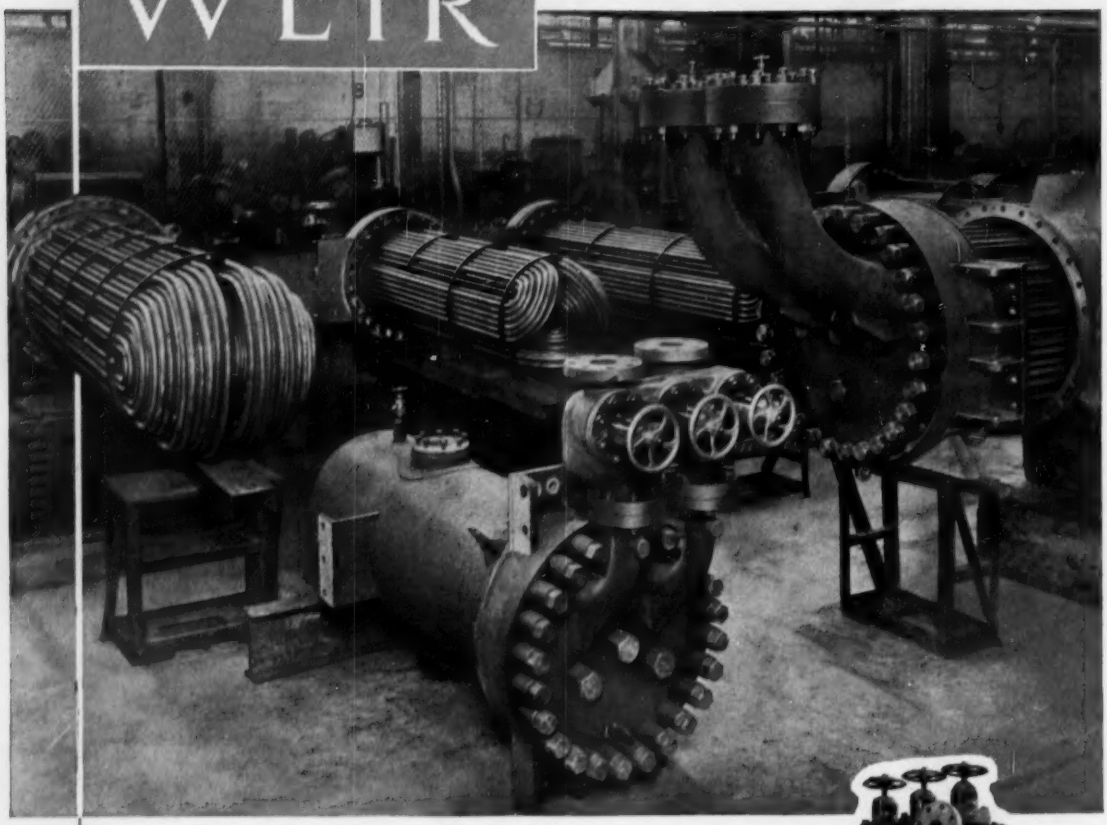

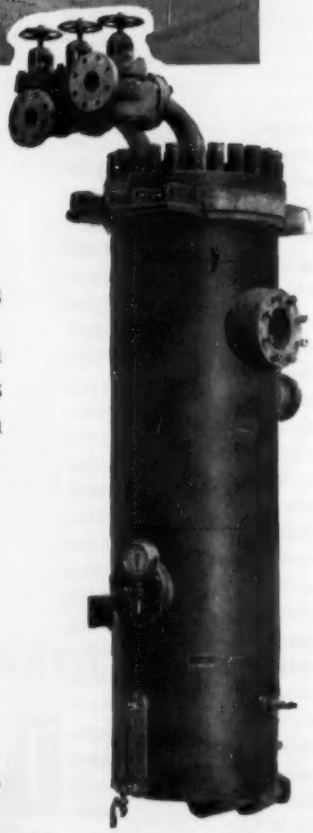

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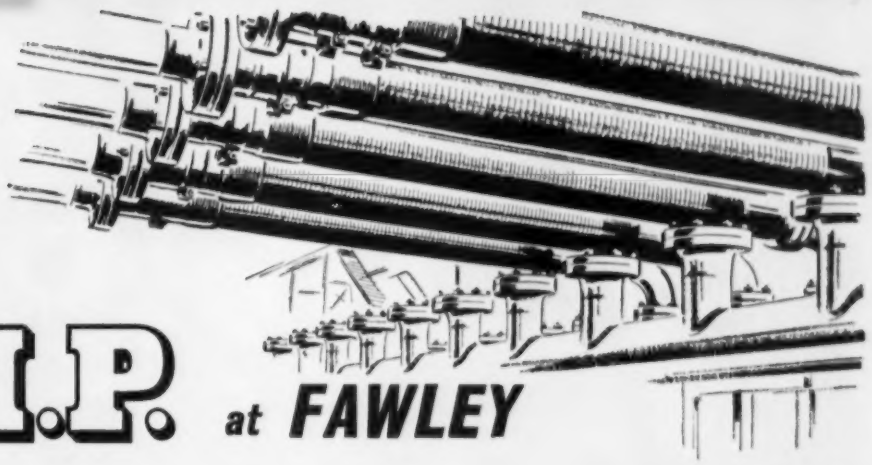
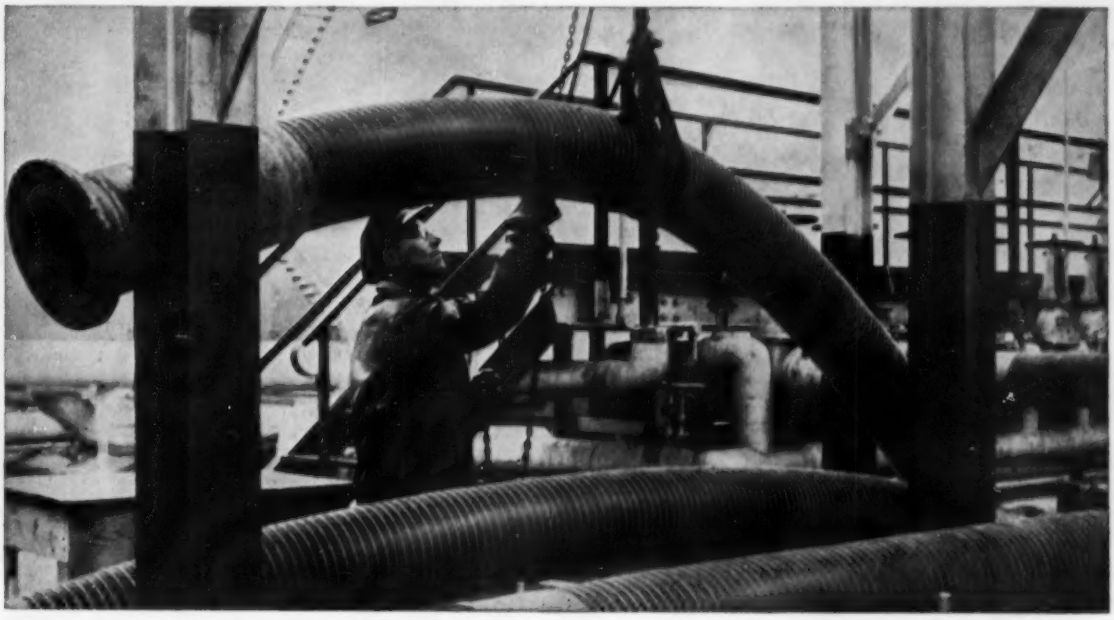


*V.I.P. at FAWLEY



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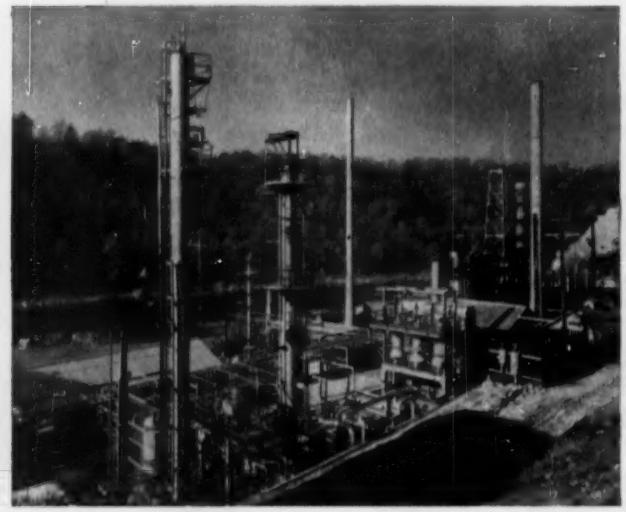
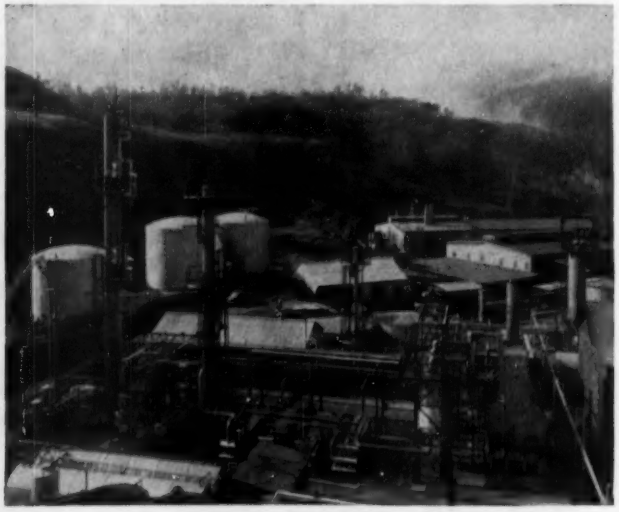
Standing out in bold relief are the reactor-heater section (left) and compressor house of the 1,500 barrel per day UOP Platforming unit — the largest of three built for Quaker State Oil Refining Corporation — at the company's Farmer's Valley, Pa., refinery.



The Allegheny River is only a stone's throw from the Platforming unit at Quaker State's refinery at Emlenton, Pa. The unit has a daily capacity of 1,000 B/SD.



Rolling hills loom behind the 1,000 barrel per day Platformer at the St. Marys (W. Va.) refinery of Quaker State. The unit is a duplicate of the one at Emlenton.

Three UOP Platforming Units Solve Pressing Octane Problem For Quaker State Oil Refining Corp.

By Guy B. Hunter
Executive Vice President
Quaker State Oil Refining Corporation
Oil City, Pennsylvania

OUR company recently concluded a three-way program of product improvement which was highlighted by the installation of three UOP Platforming units, thus giving Quaker State Oil Refining Corporation the distinction of being the first independent refiner to build three of these units at one time.

Guy B. Hunter

Quaker State for more than half a century has been world renowned for its lubricating oils. Its gasolines, distributed under the brand name of "Sterling," have enjoyed an equally high reputation in the marketing areas of our three refineries at Farmer's Valley and Emlenton, Pa., and St. Marys, W. Va. The English pound sterling sign, symbol of our gasolines, always has been the motorist's assurance that he was purchasing the highest quality motor fuels that the petroleum refining industry was able to produce.

We experienced little or no difficulty in meeting octane requirements as the trend moved upward toward more powerful gasolines, but as the level approached 93 we were confronted with a serious problem because our thermal cracking equipment could not be operated economically above that figure. It was therefore apparent that if "Sterling" gasolines were to continue their leading position, another way must be found to increase our octanes.

Planned for the Future

We not only had the immediate problem of producing a higher quality motor fuel but we, like so many other independent refiners, also had to "take out insurance" against the day when octane levels would be still higher. A thorough study of the catalytic reforming field convinced us that the UOP Platforming process was indeed the answer to the situation and would give us an ample octane potential cushion for the future.

We based our decision to install Platforming at all of our refineries on four major factors which were, briefly, the commercially proved experience of Platforming; the proved reliability and the many years of satisfactory service we received from Universal Oil Products Company; the high degree of market flexibility which the process would give to our operations, and the conviction that we would be equipped to meet all anticipated octane increases in the foreseeable future.

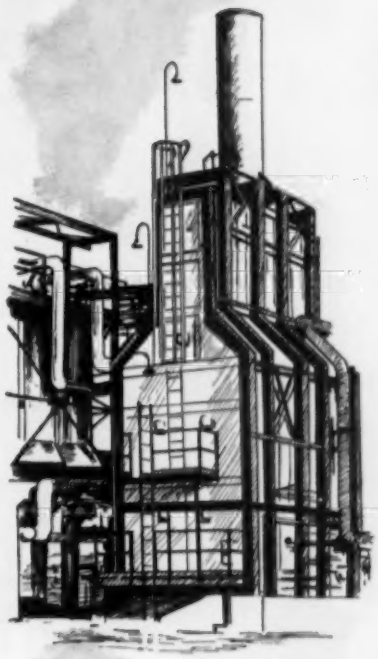
Operation of the three Platformers has convinced us that our selection was a wise one. "Sterling" gasolines now are among the highest octane motor fuels available in our marketing areas, which are concentrated through the heart of the Allegheny Mountains in Western Pennsylvania and along the upper reaches of the Ohio River in West Virginia. Additionally, through Platforming, we have been able to increase substantially our overall cracking capacity by utilizing existing thermal reforming equipment for more extensive cracking service.

Full Boiling Range Gasoline

Quaker State is among the first to produce a 98 leaded octane Platformate. Furthermore, when the unit at Farmer's Valley went on stream, it marked the first time that a Platformer in the Eastern part of the United States was operating on full boiling range gasoline.

The same type of charge stock is fed to the Platforming units at Emlenton and at St. Marys. It boils in the range of 80° F. to 400° F., and it is low in naphthene content. Despite the fact that the feed stock contains an unusually high contaminant concentration, special provisions in the design of the units have enabled us to maintain satisfactory operations at the desired octane number.

We are more than satisfied with the operation of our three Platforming units, which give us a daily capacity of high quality gasoline as follows: 1,500 B/SD at Farmer's Valley and 1,000 B/SD each at Emlenton and St. Marys. From the standpoint of economics, we are pleased to report that Platforming has made it possible for Quaker State to produce 98 leaded octane gasoline at a yield of 81.8 per cent as against our previous operations by which we produced a 93 leaded octane product at 72 per cent yield.



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Supplied ready mixed for immediate use. Suitable for rammed linings and patching existing Durax No. 1 or firebrick linings. Service temperature range 1300/1650°C.

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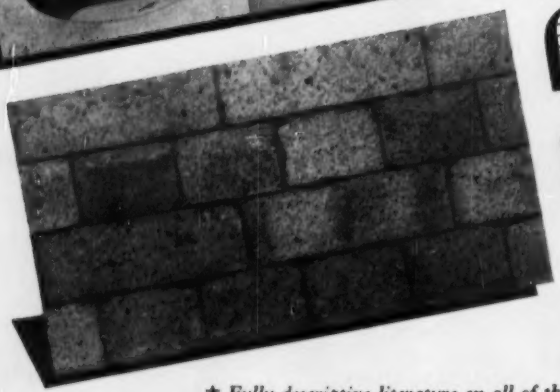
For casting in situ and making special shapes. Supplied dry. Special characteristics include—rapid setting as hard as firebrick: pours into position: no permanent volume change: little tendency to spall. Maximum service temperature 1300°C.



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A cement designed for application with a pneumatic gun. Alternatively it may be plastered into position. Durax No. 5 has little rebound loss when applied by gun to Catalytic Regenerators and forms a dense tough lining with a maximum service temperature of 1300°C.

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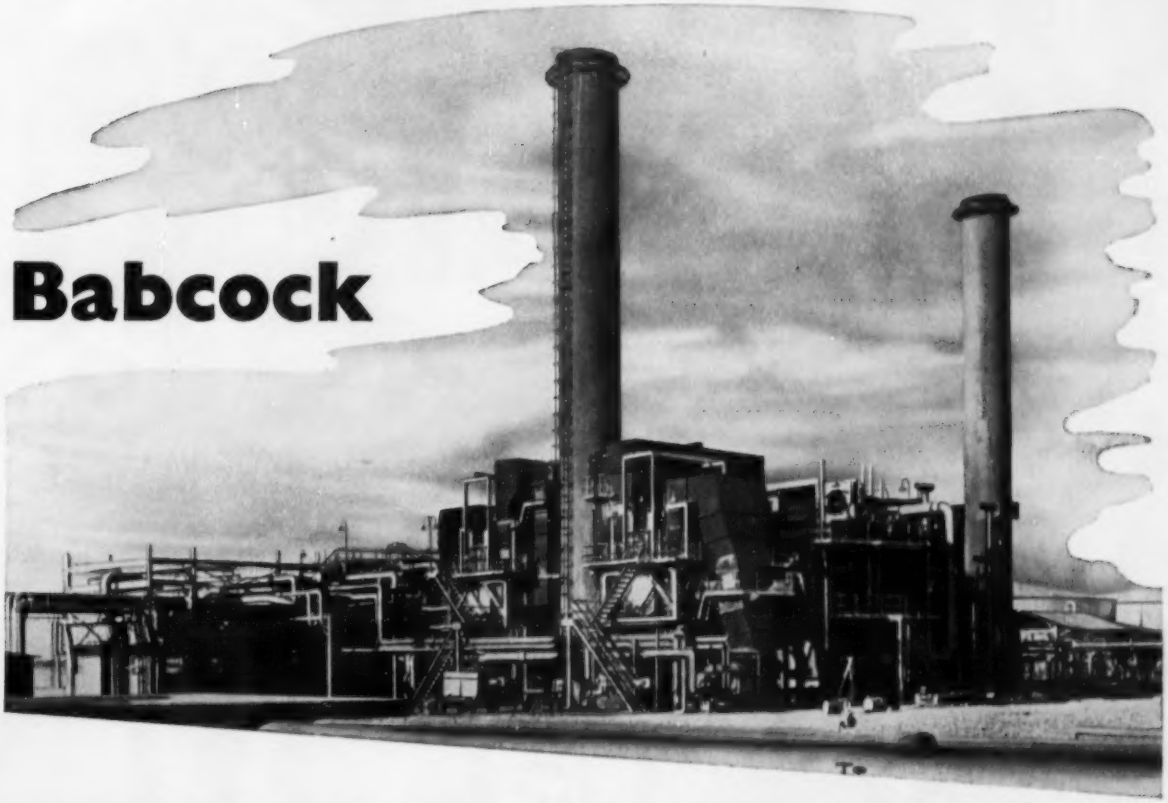
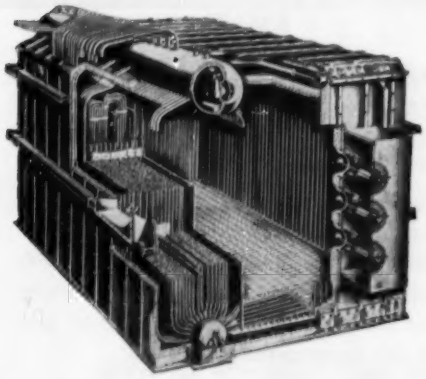


equipment for the Oil Industry

Above: *An "outdoor" installation of Babcock Integral Furnace boilers at Shell's Shell Haven Refinery, Essex, England.*

Below: *Sectional view of an oil-fired Babcock Integral Furnace boiler.*



A notable trend in the equipment of post-war oil refineries and chemical plants has been the use of "outdoor" boiler installations, in which only the firing floor and control points are under cover and no boiler house structure is involved.

The Babcock Integral Furnace Boiler*, modern in design, compact, flexible in operation, with quick steaming characteristics and high availability, has proved ideal for this and many other industrial applications. It is available in capacities from 40,000 to 250,000 lb. steam/hr., for steam conditions up to 900 lb. sq. in. and 900°F., with oil, gas or coal firing, singly or in combination.

Babcock & Wilcox Ltd. manufacture a wide range of equipment for the oil and chemical industries, including complete steam raising plants, welded pressure vessels, heat exchangers and plant for waste heat utilization.

* *Ask for Publication 1525*

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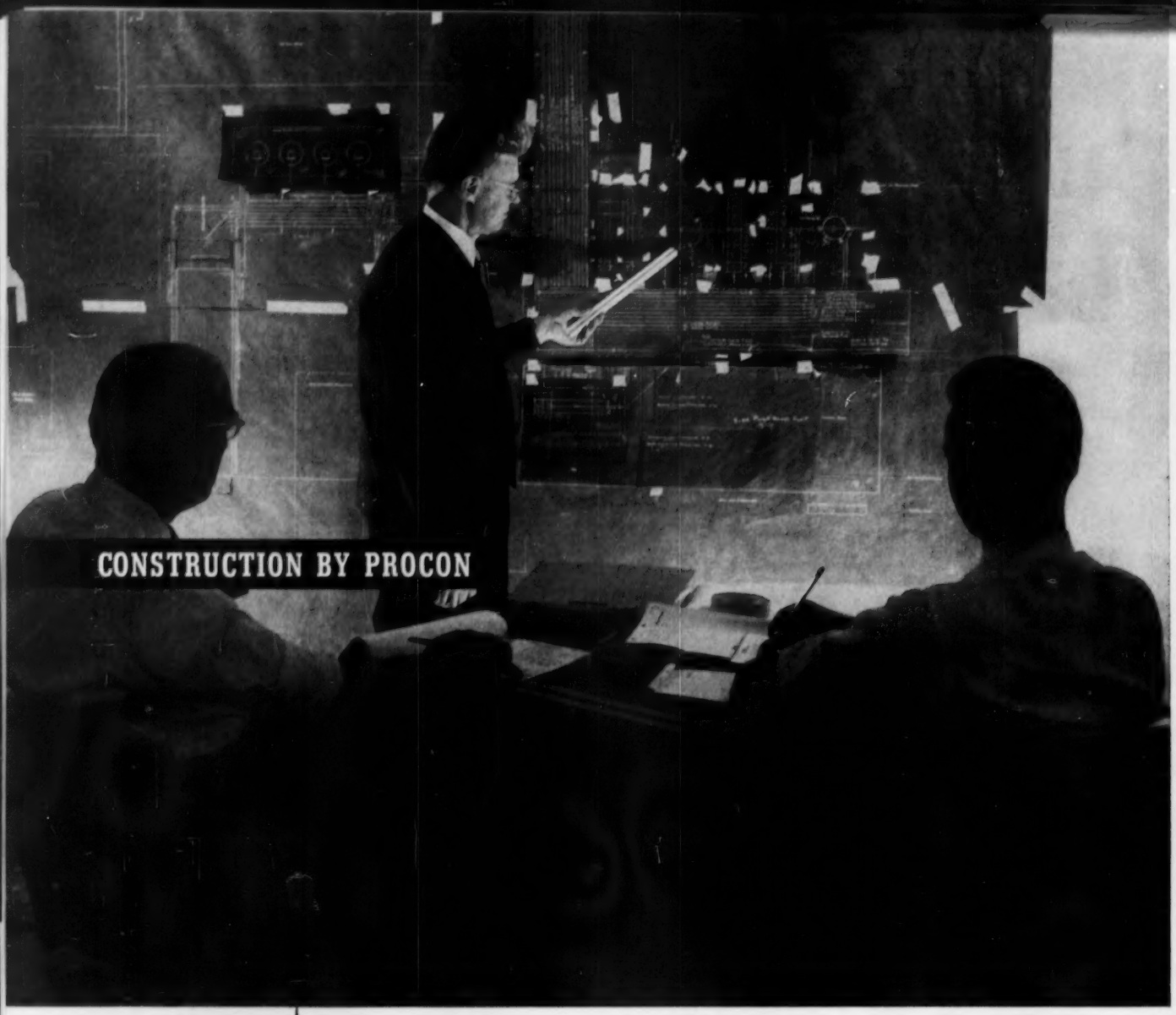

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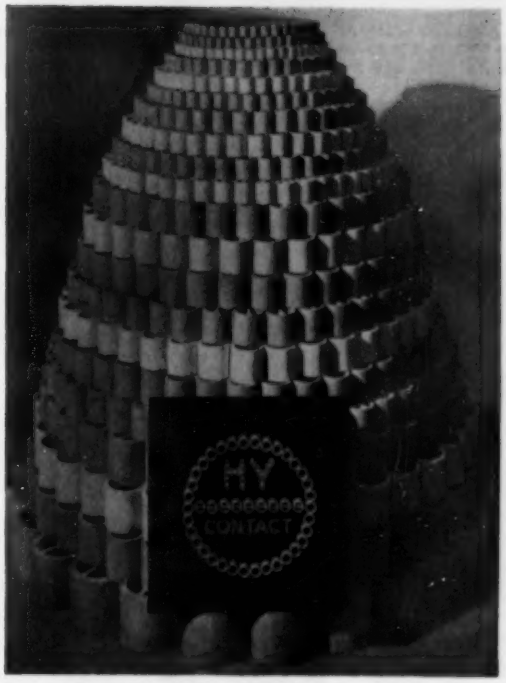

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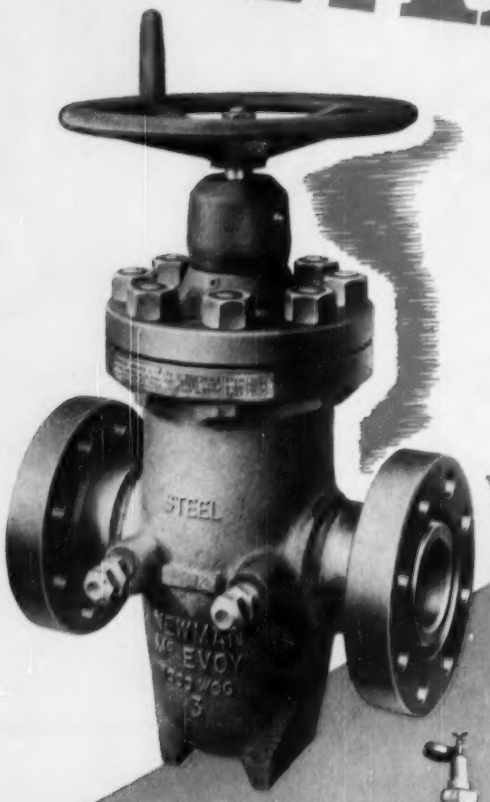
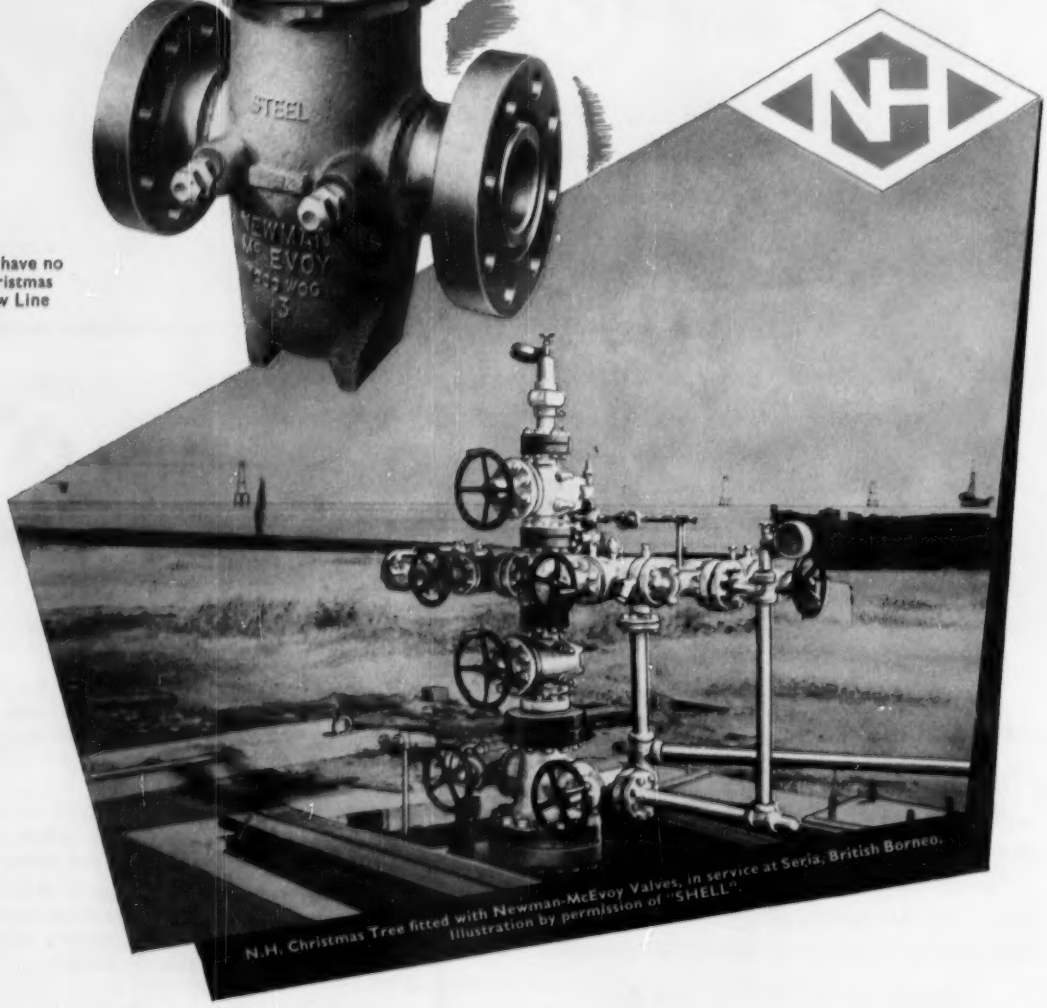


These Valves have no equal for Christmas Tree and Flow Line service.

AUTOMATIC SELF-SEALING CONDUIT GATE VALVES

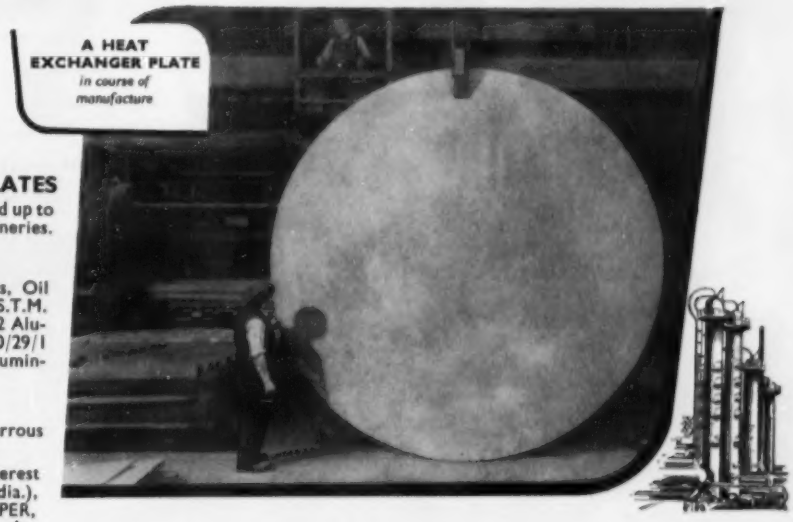
Newman–McEvoy Valves are available with Screwed, Socket Weld or Flanged Ends (illustrated). Valves for 2,000 lb., 3,000 lb., 5,000 lb., and 10,000 lb. W.O.G. pressures are manufactured in 2", 2½", 3", and 4" sizes. Write for illustrated and descriptive brochure.



N.H. Christmas Tree fitted with Newman-McEvoy Valves, in service at Seria, British Borneo. Illustration by permission of "SHELL".

Newman, Hender & Co. Ltd. WOODCHESTER STROUD GLOS.

Tubes and Plates for Heat Exchangers



A HEAT EXCHANGER PLATE *in course of manufacture*

Birmingham Battery "True to Specification" products in non-ferrous metals are widely used in the Oil industry.

"BATTERY" CONDENSER PLATES
in Naval Brass or Yellow Metal are produced up to the heaviest sizes required by the Oil Refineries.

CONDENSER TUBES
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BI-METAL TUBES
for combining the properties of Non-Ferrous Tubes with Steel Tubes.

- Other "BATTERY" manufactures of interest to the Oil Industry are TUBES (up to 24" dia.), SHEETS, STRIP, ROD and WIRE in COPPER, BRASS, PHOSPHOR-BRONZE etc., to the latest British Standard Specifications. Where necessary, we should be pleased to work to customers' own requirements.

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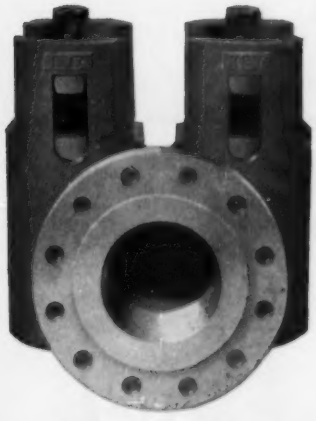

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MILLENNIUM-KEY CAST STEEL FITTINGS

for strength & pressure tightness

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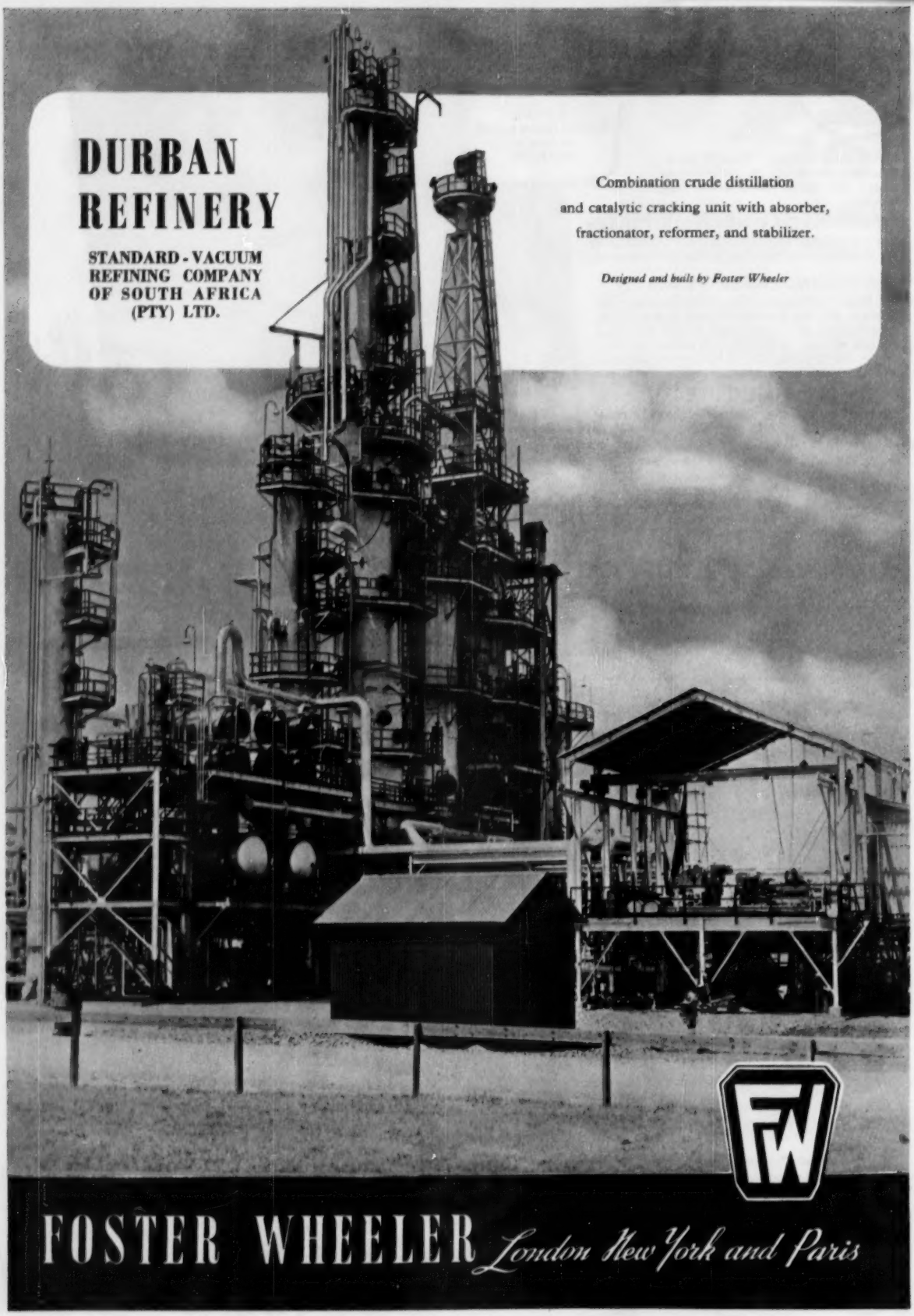
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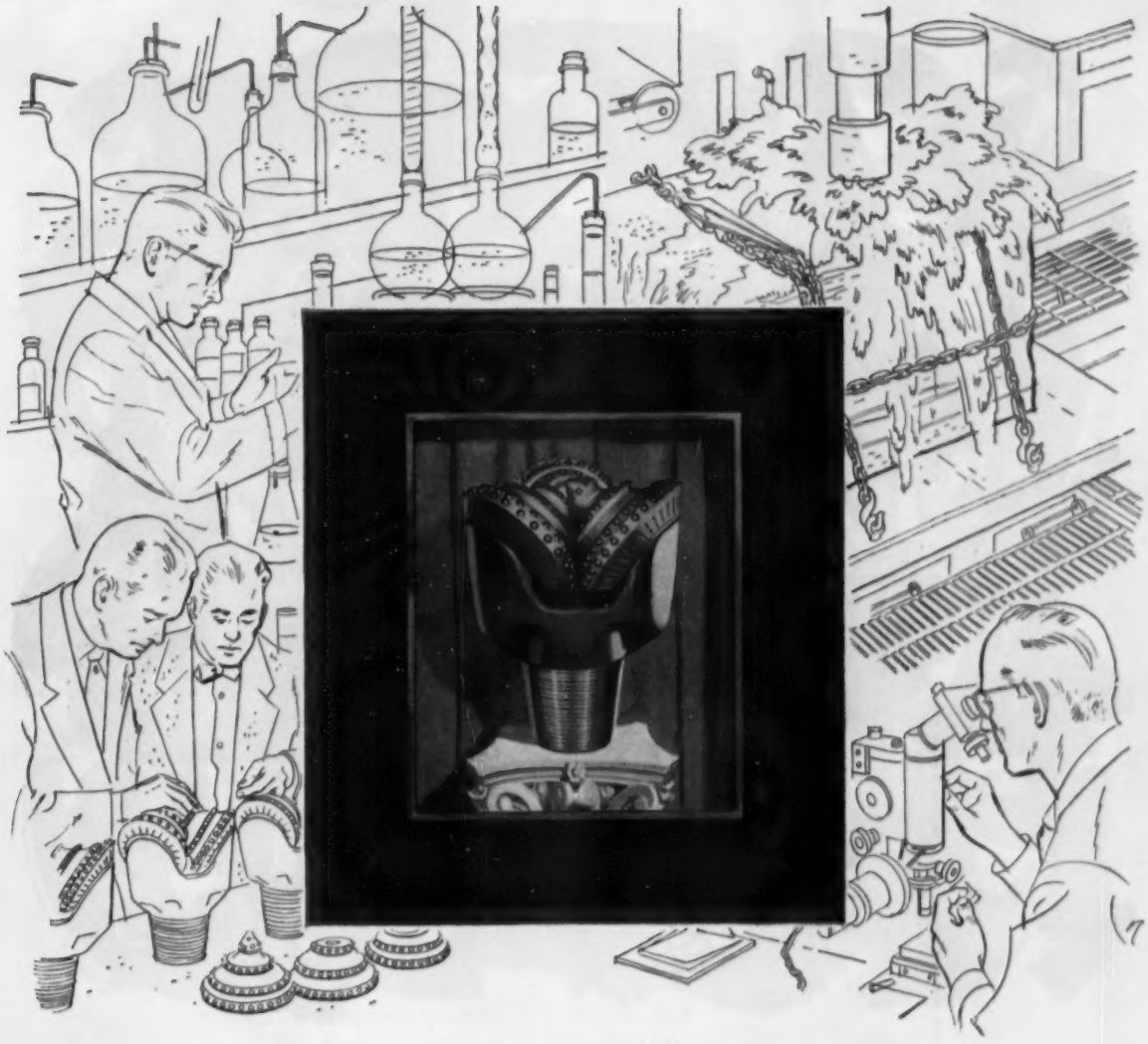

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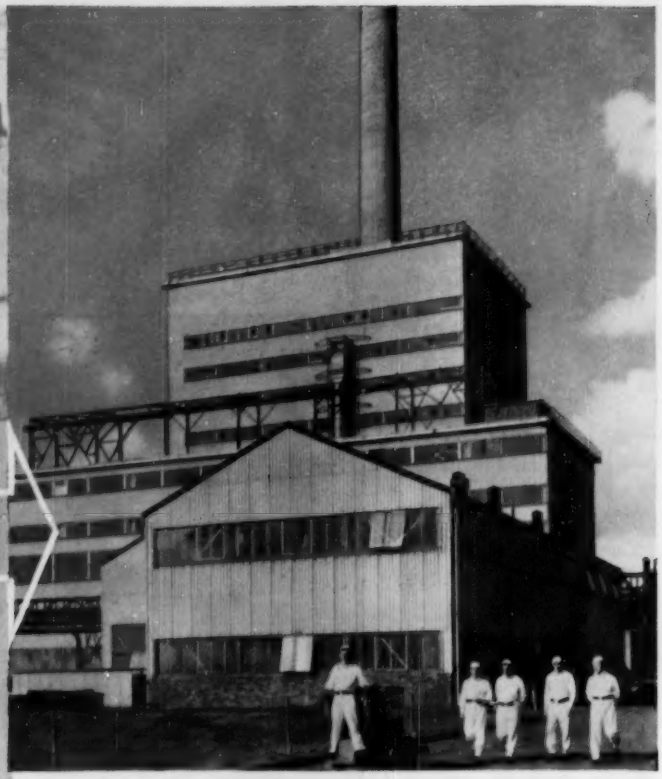

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